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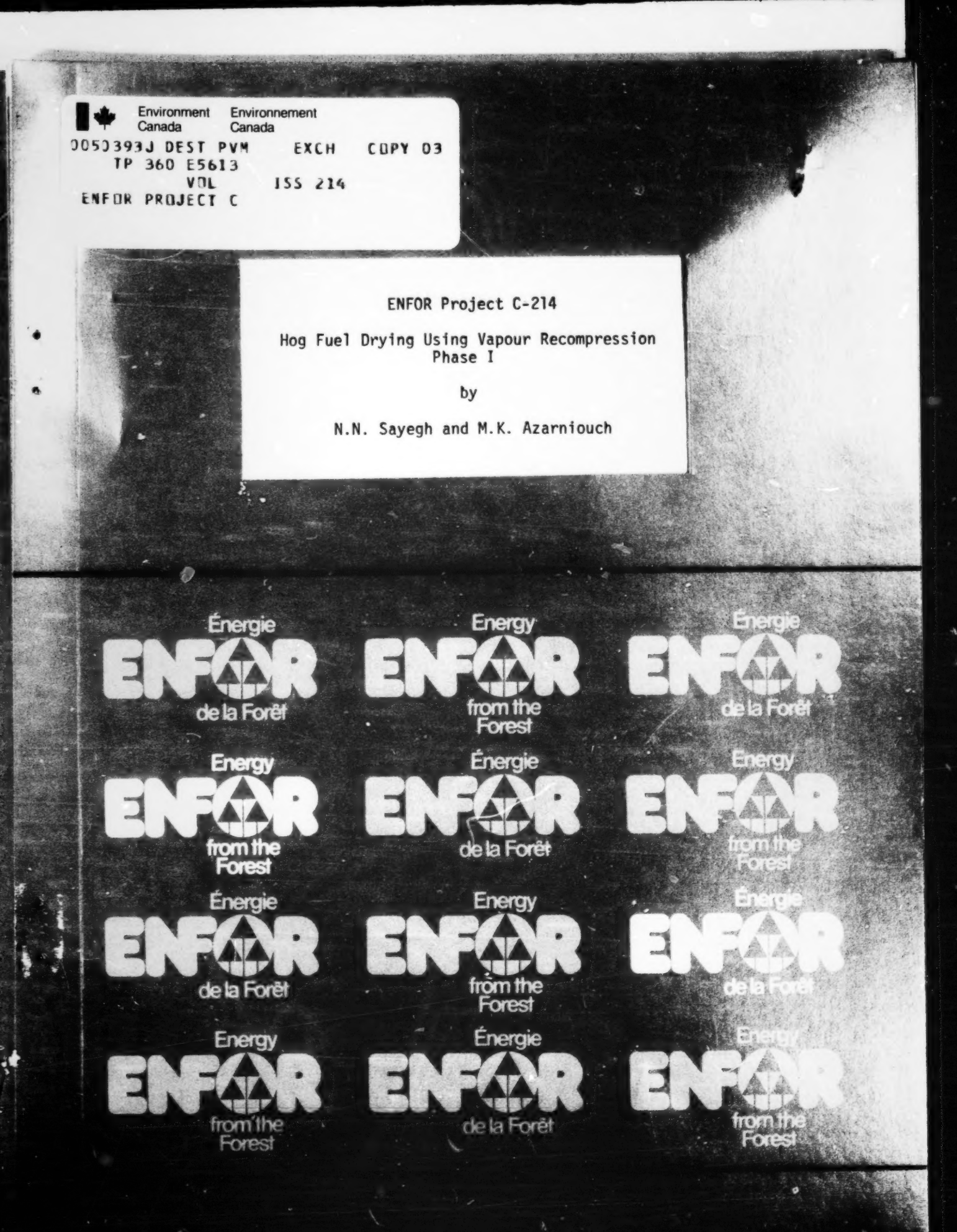
ENFOR Project C-214

Hog Fuel Drying Using Vapour Recompression
Phase I

by

N.N. Sayegh and M.K. Azarniouch



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ENFOR Project C-214

Hog Fuel Drying Using Vapour Recompression
Phase I

by

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August 1982

TEARSHEET ANNOTATION

Compared to drying hog fuel by direct contact with the heat transfer surface, increases of 2 to 6 times in the overall heat transfer coefficients were possible by the use of an intermediate (liquid) heat transfer medium.

ABSTRACT

The overall heat transfer coefficients for the drying of hog fuel were determined under conditions simulating vapour recompression, i.e. when the heat is supplied by the indirect condensation of low-pressure steam. The tests were carried out both in absence of an intermediate heat transfer medium where the contact between the solids and heat exchange surface was provided by rotating the drying vessel, and with tall oil as a heat transfer medium with relative motion provided either by again rotating the drying vessel or by circulating the oil past the solid particles in a stationary drying vessel.

The overall heat transfer coefficient decreased rapidly with the decreasing moisture content. In the tests without the oil the average overall heat transfer coefficient between 20 and 60% moisture was 55 W/m^2K. The average heat transfer coefficients obtained with the stirred-oil tests were somewhat higher, i.e. between 90 and 150 W/m^2K. The highest coefficients ($300W/m^2K$) were obtained when the oil was circulated through the dryer. For this reason, the next phase of this study will comprise the testing of a continuous pilot-scale equipment where the oil/solids mixture will be made to flow co-currently through the heat exchanger.

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HOG FUEL DRYING USING VAPOUR RECOMPRESSION

FINAL REPORT

INTRODUCTION

Hog fuel is a broad term used by the forest products industry to describe all types of wood residues that are used as fuel. These can be bark, wood chip rejects, sawdust, shavings, etc. The moisture content of these residuals depends mainly on their source and on the conditions of storage. Typically, shavings and sawdust obtained from the lumber industry have low moisture contents. The moisture content of bark from dry debarkers is similar to that of uncut wood, namely, about 50%*. On the other hand, wet debarkers which use water in their operation produce bark with moisture contents exceeding 70%, but this is usually reduced to about 55-65% by pressing. The moisture content of hog fuel stored in outdoor piles is influenced by two factors: (i) the material on the surface of the pile can absorb or lose moisture depending on the weather conditions, and (ii) heat generation within the pile, due to some degradation reactions, can cause drying of the inner material.

The introduction of the significant amounts of water contained in the hog fuel impairs the efficiency of operating a hog-fuel boiler by: (i) reducing the net heating value of the fuel, (ii) increasing the particulate emissions caused by incomplete combustion, (iii) requiring higher air flows to maintain combustion, (iv) decreasing the rate of combustion, and (v) reducing the heat flux as a result of lower flame temperatures. One study [1] carried out on a pulp mill boiler, has shown that the steam output could be increased from 27,000 to 50,000 kg/h when hog fuel at 63% moisture was dried to 28% moisture prior to

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* All moisture contents in this report are given on wet basis.

combustion. In this study, the oxygen content was maintained at the minimum allowable level for the system to operate without smoke, while the amount of hog fuel used was determined by its rate of combustion. Therefore, significant improvements in the energy efficiency and the steam producing capacity of hog-fuel boilers could be achieved by the use of an economical hog-fuel drying system.

Hog-Fuel Dryers

Several systems are commercially available for the drying of hog fuel and most of them use the waste heat in the flue gas as the source of heat for the drying operation. Commercial flue-gas dryers are rotary, cascade or flash type. However, all of these suffer from the same inherent disadvantage:

(i) Only a limited fraction of the heat in the flue gas can be utilized for drying, since the moisture introduced with the hog fuel as water leaves the system as vapour carrying with it a significant amount of unrecoverable energy as latent heat.

(ii) If higher temperature or a large flow of flue gas is available, over-drying can cause distillation and loss of combustible volatile matter with the flue gas, as well as excessive dusting. In order to avoid this, the average moisture content of the dried fuel is usually kept above 35 to 40% [2].

(iii) There is a danger of fire and explosion which, in some instances, have occured as a result of contact between the over-dried fines and the oxygen-containing hot flue gas.

In a different type of dryer [3,4], finely comminuted bark is dried while suspended in recirculating superheated steam; the heat is supplied indirectly by the condensation of high-pressure (1 MPa) steam extracted from a back-pressure turbine. The operation is kept air-free and under pressure to allow the recovery of the evaporated moisture as 'dirty' steam at intermediate pressure (about 500 kPa), which can be used either indirectly to generate clean steam at lower pressure, or directly in such applications as chip pre-heating. The disadvantages

of this scheme are that it can be appplied only in mills which practice congeneration, and that full credit must be given in using the generated dirty steam in order to make the process economical. Also, this process requires comminution of the wet hog fuel feed which, in addition to its cost, would result in a finely-divided product that is difficult to handle and store.

For a more comprehensive comparison of commercial hog-fuel dryers, including economic assessment, the reader is referred to the ENFOR report on Project C-96 [5].

Vapour Recompression Drying

In a vapour recompression operation, the vapours produced in the drying or evaporation process are compressed and returned back to the unit to supply, by indirect condensation, most of the heat requirements of the operation (Figure 1). In such a system, the heat is transferred from the condensing compressed vapours, across the heat exchange surface to the material to be dried or concentrated. The driving force for heat transfer is provided by the higher condensation temperature of the compressed vapours. Because of economic considerations, the compression ratios in such systems are, typically, kept less than two, leading to overall temperature differences of less than 20°C. However, lower temperature difference means an increase in the heat transfer surface area requirement. Usually, the optimum design of a vapour recompression system is obtained by minimizing the total costs of the compressor and the heat exchanger.

The net energy savings with a vapour recompression system are typically high, because the moisture entering with the wet material as water leaves the system also as water , although at a higher temperature and theoretically, no energy would be lost as latent heat. The main energy input to a vapour recompression system is the electricity required to operate the compressor, while some additional energy, supplied by the external steam or possibly flue gas, may also be required to meet the total energy demand of the system.

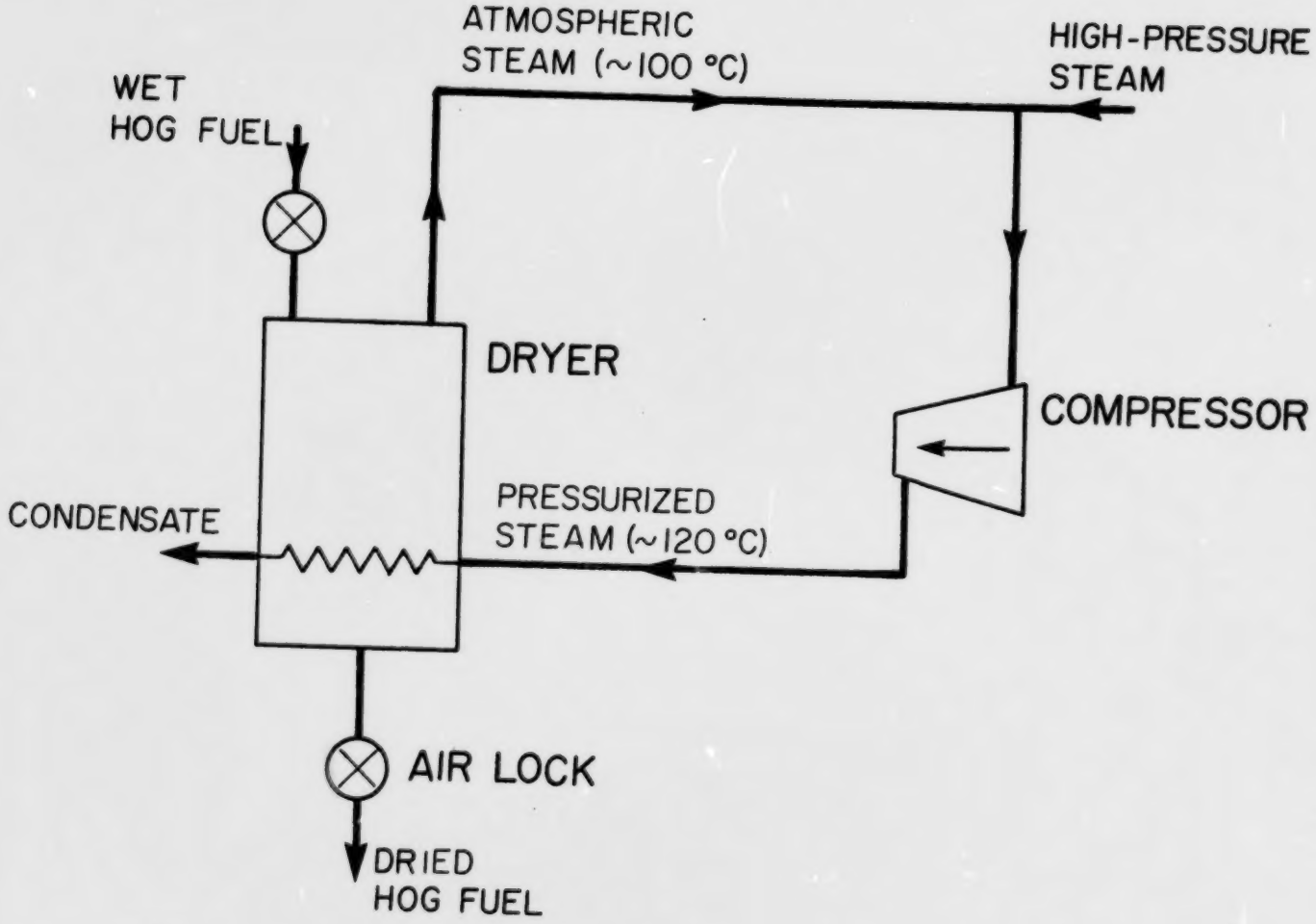


Figure 1. Conceptual scheme of a vapour recompression dryer.

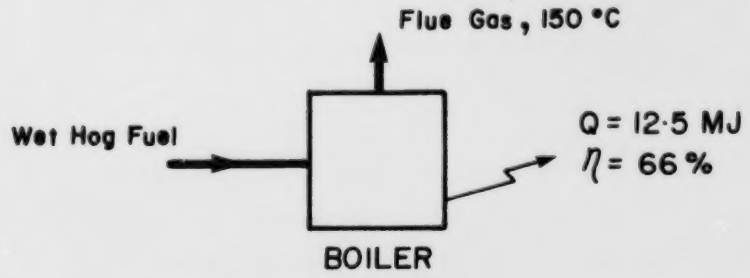
Figure 2 shows a comparison of the overall theoretical energy efficiencies of burning hog fuel with and without a dryer. The assumption made in the calculations were: (i) the initial hog-fuel moisture content is 60%, (ii) the gases leave the boiler at 150°C and (iii) 30% excess air is used in the boiler. When no dryer is used, the amount of heat available for raising steam is 12.5 MJ/kg fuel (dry basis), i.e. only 66% of the gross heating value of hog fuel. Installing a flue gas dryer with gas exit temperature of 100°C would reduce the moisture content of the fuel entering the boiler and, consequently, increase the available heat to 13.1 MJ/kg. If a vapour recompression dryer, designed to reduce the moisture content to 20%, is used, then the net heating value of the fuel (after substracting the steam requirements of the dryer) would increase to 15.4 MJ/kg. The electricity required to operate the compressor would be 0.3 MJ/kg fuel. In other words, a 2.9 MJ/kg fuel increase in the net heating value of the fuel could be achieved by the use of only 0.3 MJ/kg of external energy, i.e. the operation has a coefficient of performance of about 10. Details of the calculations are given in Appendices I and II.

Proposed Process

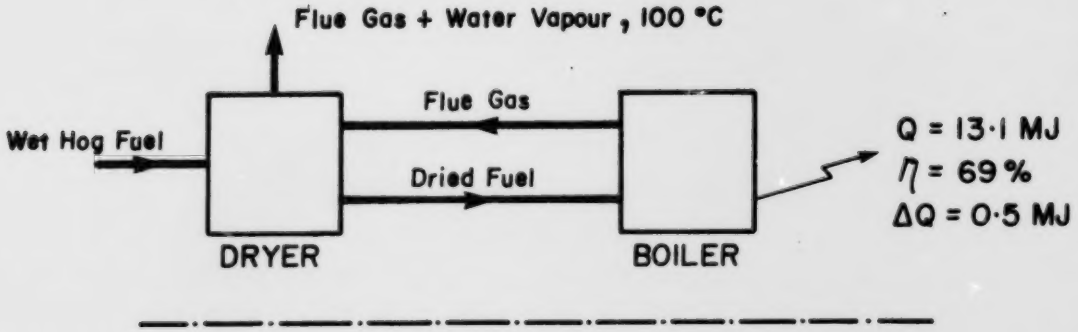
The hog-fuel drying process proposed here is based on the principle of vapour recompression described above. The wet hog fuel would be fed into the dryer vessel through an air lock to minimize the amount of non-condensables (e.g. air) in the vapours leaving the dryer. The size of the dryer and the heat transfer surface would be designed to provide sufficient residence time and energy for the hog-fuel particles to reach the desired degree of drying. The dried particles would then be drawn out of the dryer through another air lock. Because of the absence of flue gas or air flows, the heat has to be transferred directly from the heating surface to the hog fuel particles. For satisfactory operation of such systems, careful consideration must be given to the design of the dryer in order to provide the highest heat transfer rates, both by conduction and convection, and to allow easy passage of the hog fuel through the dryer. The vapours leaving the dryer would be superheated either directly by the injection of some high-pressure

BASIS : 1 kg (dry) hog fuel ; 60 % moisture
calorific value = 19 MJ/kg (dry)

OPERATION WITHOUT DRYER



FLUE GAS DRYER



VAPOUR RECOMPRESSION DRYER

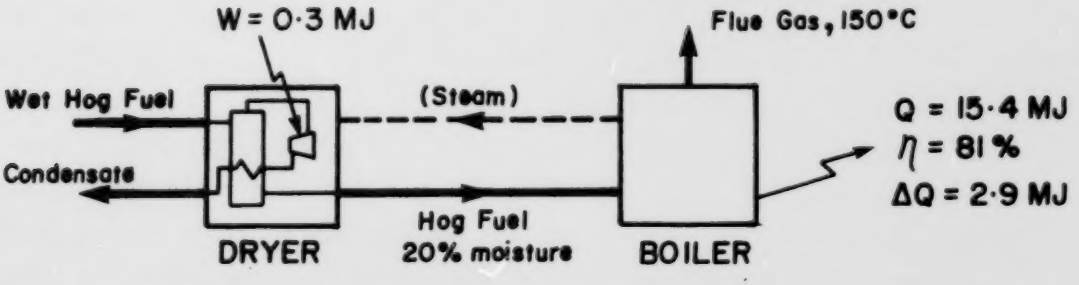


Figure 2. Comparison of the theoretical energy efficiencies of different hog-fuel dryer/boiler systems.

steam or indirectly by boiler flue gas, and compressed before introduction into the heat exchange tubes. The pre-heating step is necessary to avoid any condensation in the compressor and it also helps to meet the total energy demand of the system.

To improve heat transfer in the dryer, a combustible, non-volatile and water-immiscible liquid heat transfer medium can be used, in which the hog-fuel particles would be suspended; this liquid would be heated indirectly by the condensing steam. Since the liquid can provide a more intimate contact with the hog-fuel particles, higher heat transfer rates can be expected, leading to smaller and cheaper drying units. For such a system, a scheme like the one shown in Figure 3 might be used. The wet hog fuel would enter the dryer mixed with the recirculating oil, and the dried particles would be separated from the recirculating oil in a screw press. The location of the points of introduction and withdrawal of the hog fuel from the dryer vessel would depend largely on the density of the solids before and after drying relative to that of the hot oil.

The dried hog-fuel particles, which may still contain an appreciable amount of oil, could, if desired, be passed throuh a standard densification equipment such as a pelletizer, extruder, etc. to form a dense fuel of high heating value. In the densifiction step, additional oil containing some wood extractives would be released; this oil would be recycled back to the dryer to reduce the fresh oil make-up.

The heat transfer liquid can be an oil from almost any source. However, because of its availability in kraft mills, and with the aim of eliminating the reliance on fossil fuels, the use of crude tall oil is one practical possibility. Tall oil is a purely wood-based oil and is a by-product of alkaline pulping. The recoverable amount of crude tall oil per tonne of pulp is about 20 to 40 kg depending on the wood species, while the hog-fuel yield is around 300 kg/t pulp. Therefore, oil entrainment with hog fuel should not exceed about 10%, otherwise an auxiliary source of oil would be required. However, it is expected that enough oil would be extracted from the hog fuel during pressing and compaction that only a small amount of make-up would be required.

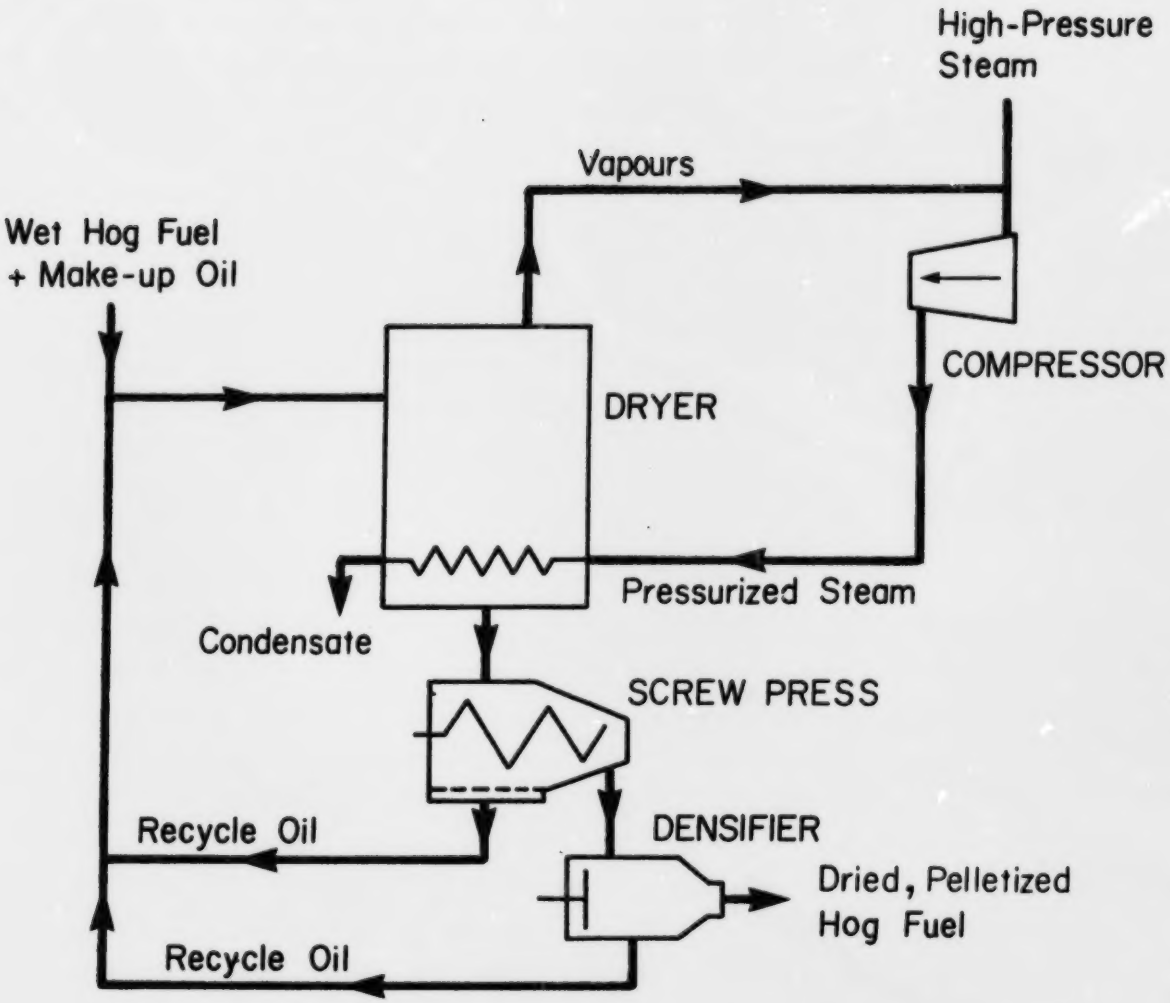


Figure 3. Schematic diagram of a vapour recompression hog-fuel dryer using oil for improved heat transfer.

OBJECTIVES

The major unknown required for the design of the vapour recompression dryer is the overall heat transfer coefficient between the condensing compressed vapour (steam) and the individual hog-fuel particles. Therefore the objectives of the experimental program were to evaluate the drying rates of hog fuel as function of (i) hog-fuel moisture content, (ii) the compression ratio and (iii) type of wood waste. A preliminary economic assessment would be preformed based on the experimental evidence and a preliminary process design with corresponding capital and operating costs.

EXPERIMENTAL

The experimental program consisted of two parts: (i) drying tests in the absence of a heat transfer medium, and (ii) drying with tall oil as the intermediate heat transfer fluid.

Drying by Direct Contact

The experimental apparatus used for tests in the absence of a heat transfer medium is shown in Figure 4. It consited of:

(i) Rotary batch dryer,
(ii) Steam generator,
(iii) Vapour condenser,
(iv) Condensate collector.

The dryer had a shell-and-tube configuration with a heat transfer area of about 0.35 m^2 (Figure 5), and it was placed on two rollers which were rotated by a variable-speed motor. Two ports in the shell allowed the introduction and the removal of the hog fuel from the batch dryer. Steam under pressure entered at one end of the tube bundle and the condensate was removed through a syphon out of the other end. The steam and condensate lines were attached to the dryer by rotary joint. The evaporated moisture left through a tube placed co-axially along the

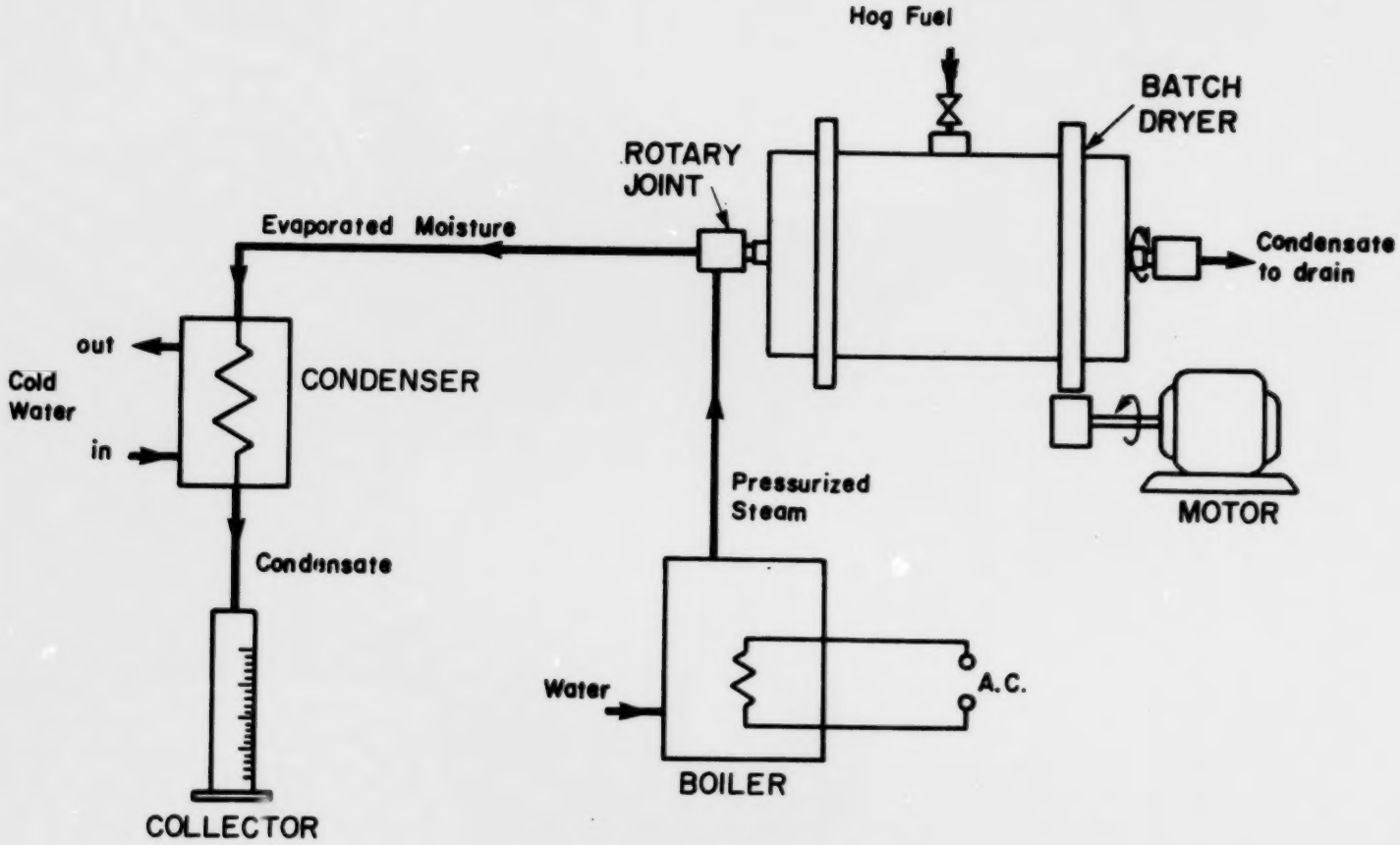


Figure 4. Schematic diagram of the experimental set-up when no oil was used as heat transfer medium.

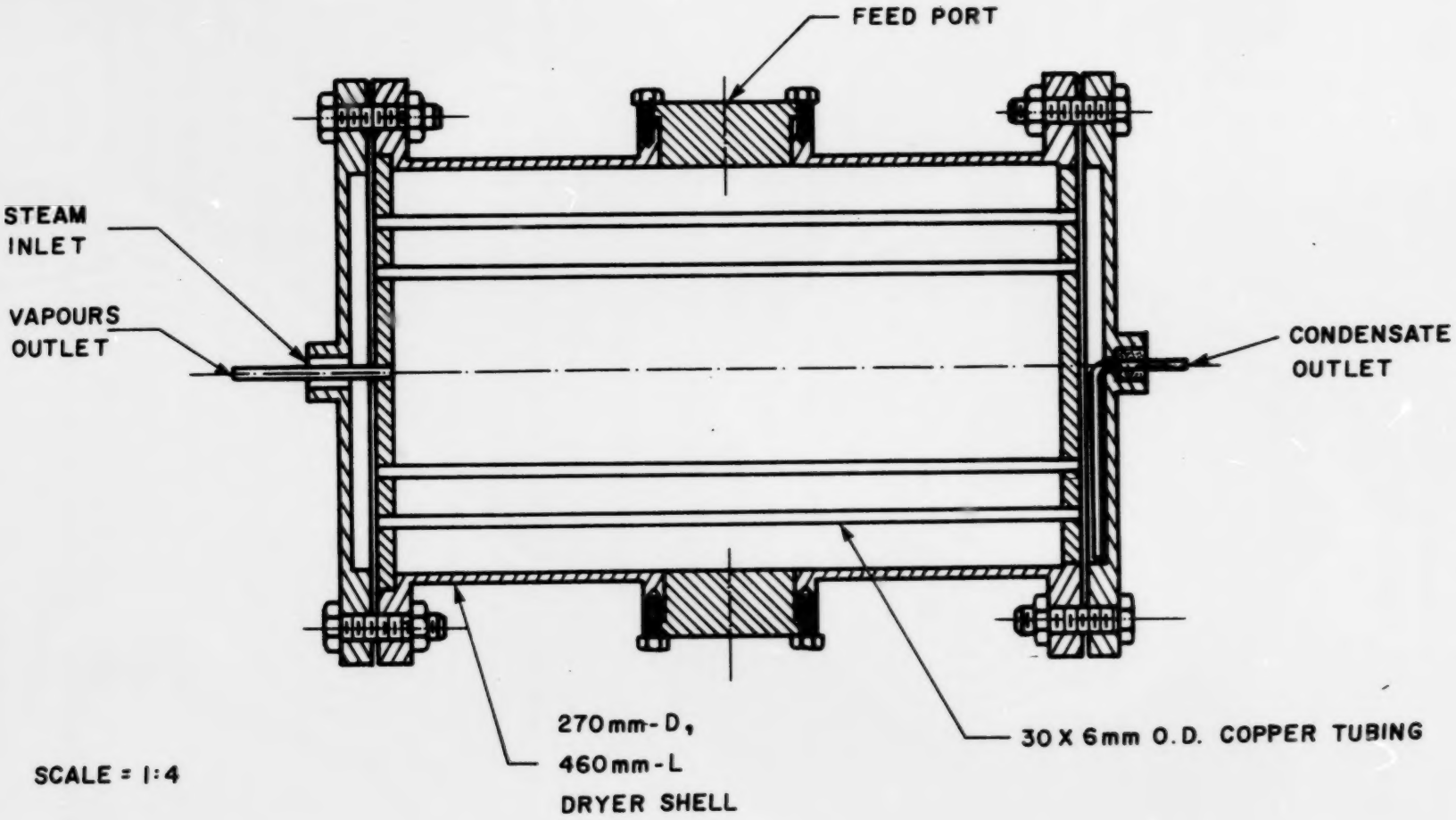


Figure 5. Details of the batch dryer vessel for tests with no oil as heat transfer medium.

steam inlet line. The vessel was rotated at about 10 rpm to provide mixing and contact between the hog-fuel particles and the surfaces of the heating tubes.

In lieu of a steam compressor, clean pressurized steam from a small electric boiler was supplied to the dryer, and the evaporated moisture was collected only for the purpose of determining the drying rates. Although vapour recompression was not applied in this study, its most important aspect, namely, the provision of the heat requirements through the indirect condensation of pressurized steam was truly simulated.

The experimental procedure consisted of charging the rotary dryer with hog fuel of known mass and moisture content. After closing the feed port, the vessel was rotated and steam from the boiler was introduced into the heating tubes. The pressure of the heating steam was maintained at 200-250 kPa with a pressure regulator in the steam line and a steam trap in the condensate line. The evaporated moisture leaving the vessel was condensed and its volume measured at fixed time intervals to give drying rates as well as intermediate hog fuel moisture contents.

At the end of each test, the heating steam was turned off and the system was allowed to cool down overnight before discharging. Opening the vessel while still hot would result in flashing off of some moisture which would upset the mass balance calculations. The final moisture content of the hog fuel was also determined to confirm the accuracy of the mass balance.

The operating variables studied were: (i) the heating steam pressure, (ii) the fraction of the dryer volume occupied by the hog fuel (degree of fill), and (iii) the heat transfer area. The heat transfer surface area was reduced by blocking half of the heating tubes; this resulted in only one-third reduction in the total surface area, since the two tube-sheets also acted as heat transfer surfaces. It was not possible to study the effect of particle size since large particles

(>40 mm nominal length) could not move freely between the tubes, and the fines (< 1 mm) tended to escape with the vapours, resulting in blockage of the condenser lines.

Drying by Stirring in Oil

In tests using tall oil for enhanced heat transfer rates, the dryer system underwent few modifications. To ensure continuous submergence of the heating surface in the tall oil, as the vessel rotated, the heating tubes were replaced with a single coil suspended just below the axis of the vessel (Figure 6). Six baffles, 20 mm wide and extending along the entire length of the vessel, were welded on the inside wall of the dryer to aid in the mixing. The stirring of the oil-hog-fuel mixture was accomplished by rotating the vessel.

An intermittent-sampling procedure was used where the vessel was rotated for 4 to 8 min., while allowing the pressure to build up inside the vessel. During sampling, the rotation was stopped and the vapours were allowed to pass through the condenser by opening the sampling port valve. Complete removal of the accumulated vapour from the dryer usually took about 3 minutes. The condenser was attached to the dryer only during the sampling period. Figure 7 shows the experimental set-up.

Drying by Circulating Oil

This work was carried out to investigate the effect of oil circulation past the heat exchange surface, on the drying rates of hog fuel. A schematic diagram of the experimental set-up is shown in Figure 8. The dryer consisted of an externally-heated vessel, 150 mm in diameter and 1 m high. The oil entered at the bottom of the vessel and left with the vapour from the top through a screen. The vapour was separated from the oil in a second tank and was then passed to the condenser and collector. The hot oil was circulated back to the dryer by means of a centrifugal pump. Heat was supplied to the dryer through the condensation of pressurized steam in the heating jacket.

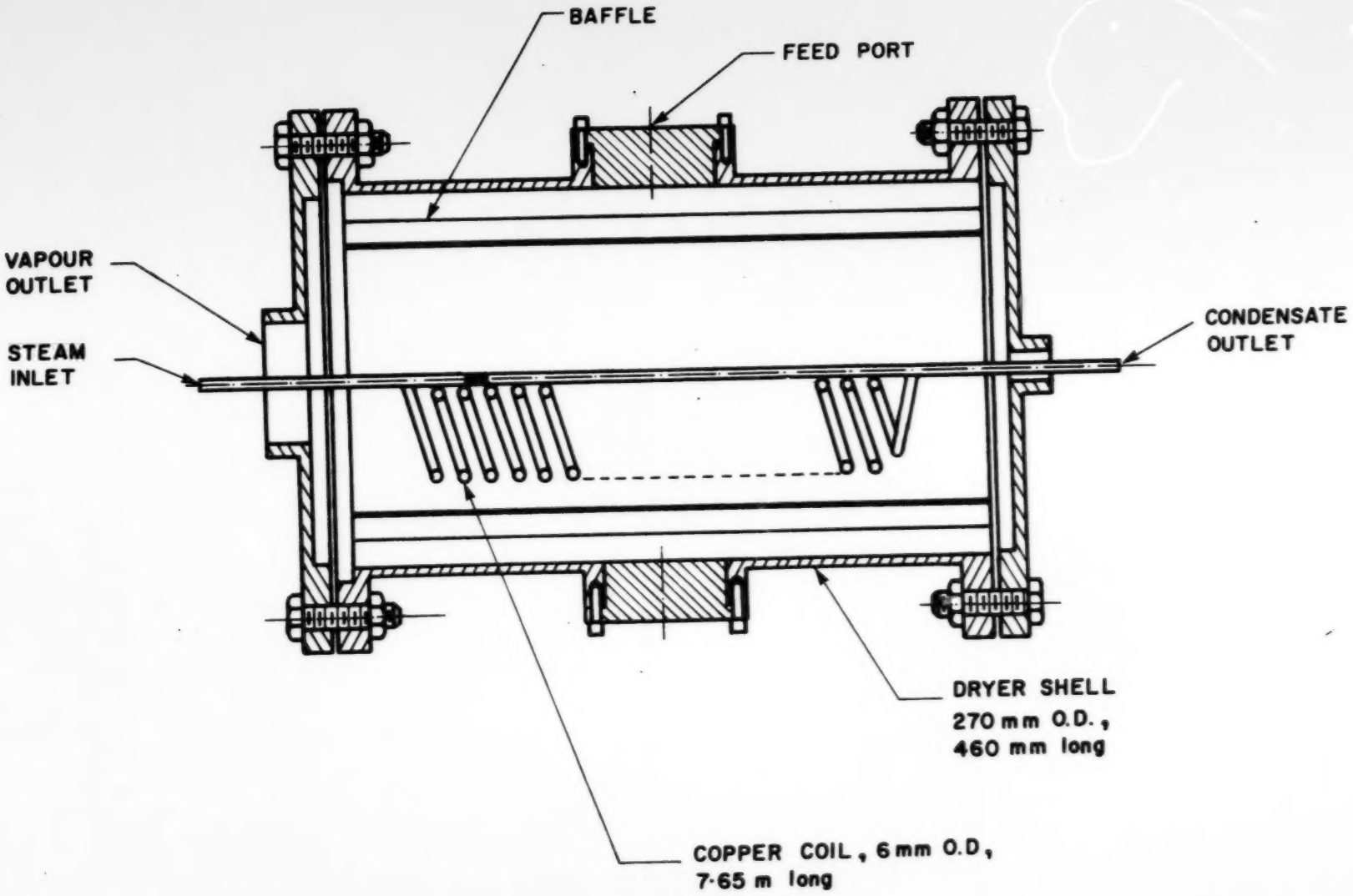


Figure 6. Details of the dryer vessel used in the stirred-oil tests.

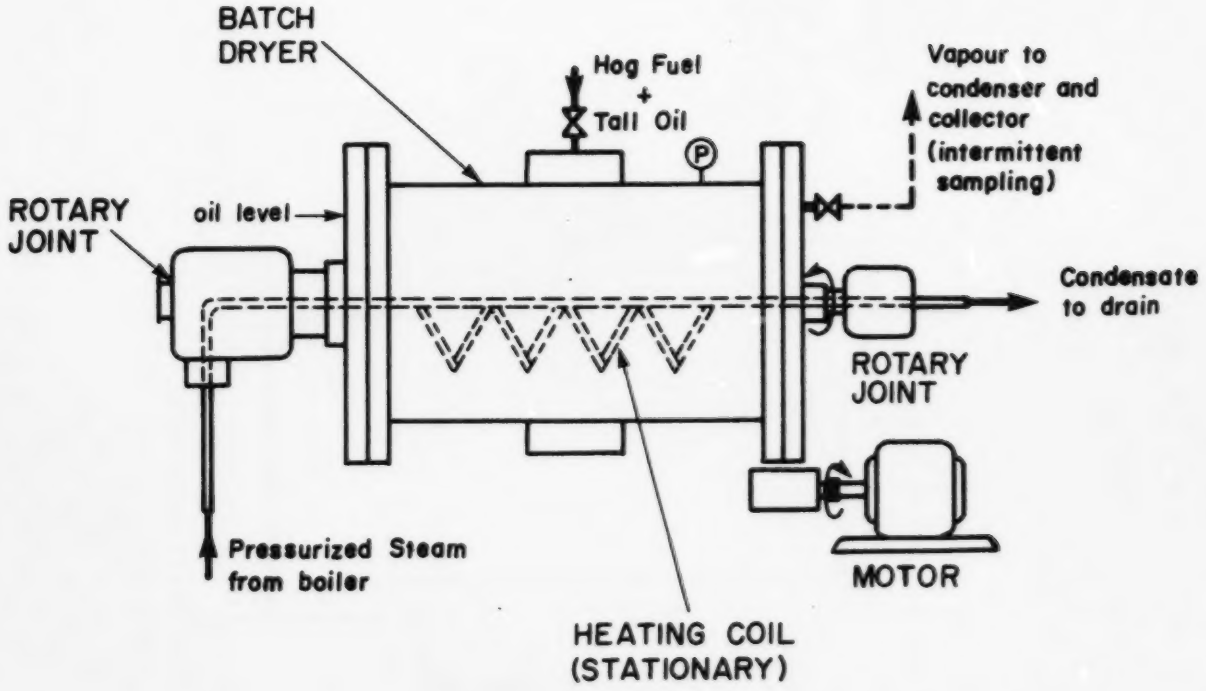


Figure 7. Experimental set-up in the stirred oil tests, with intermittent rotation and sampling.

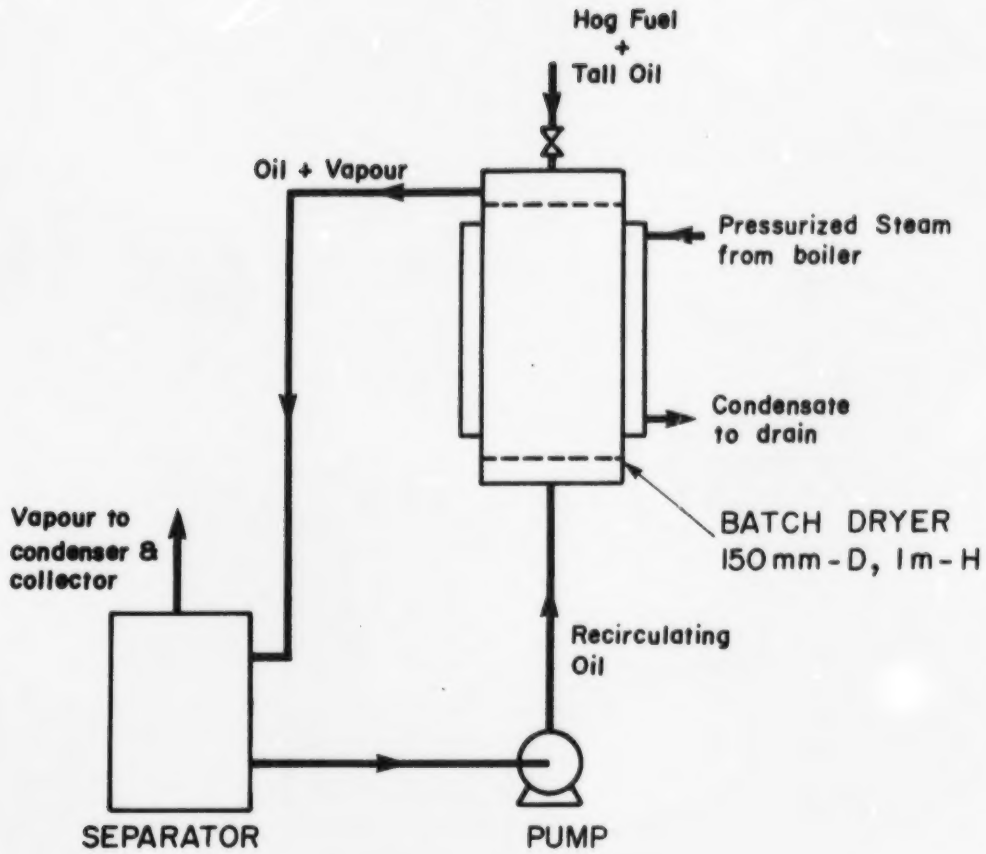


Figure 8. Experimental set-up used in the circulating oil tests.

The experimental procedure consisted of charging the dryer vessel with the hog fuel and after all the ports had been closed, starting the pump and introducing the steam into the jacket of the drying vessel. The evaporated moisture was condensed and collected at fixed time intervals to give the drying rates and the intermediate moisture contents.

One major problem encountered with this approach was the tendency of hog fuel to pack near the exit of the dryer, which led to: (i) significant reduction in the oil circulation rates and (ii) pressure build-up inside the dryer which in turn prevented evaporation of the hog-fuel moisture. Such difficulties were not encountered when wood chips instead of hog fuel were used in the drying tests. This is because wood chips are usually more homogeneous and rigid and allow a higher degree of voidage upon packing.

RESULTS AND DISCUSSION

The primary data from all drying tests were collected in the form of total accumulated condensate volume versus time. The method for converting these measurements to drying rates or overall heat transfer coefficients as function of the moisture content is given below.

Assume: initial mass of wet solids = S (kg)
initial moisture content = M
condensate mass collected after time t = W_1 (kg)
condensate mass collected after time (t + Δt) = W_2 (kg)

The moisture content at time t:

$$x_1 = (SM - W_1)/(S - W_1)$$

and, the moisture content at time (t + Δt):

$$x_2 = (SM - W_2)/(S - W_2)$$

The average moisture content during interval Δt:

$$x_{12} = 0.5\ (x_1 + x_2)$$

and the drying rate during Δt:

$$m = (W_2 - W_1)/\Delta t \qquad \text{(kg/s or kg/h)}$$

The overall heat transfer coefficient is calculated from the drying rate, thus:

$$h = m\lambda/A(T_{steam} - T_{solids}) \qquad (W/m^2\ K)$$

where λ = latent heat of vaporization of water at 100°C (2260 kJ/kg)
A = overall heat transfer area, m^2
T_{solids} is assumed to be 100°C

Since the drying rates varied with the moisture content, it was necessary to calculate an average drying rate value which would be applicable for a known range of moisture contents. This was carried out as follows:

At any moment the drying rate is given by:

$$m = dy/dt$$

or,

$$dt = dy/m$$

where y = moisture content on dry basis, kg water/kg dry solids.

The total time, θ, required to go from moisture content Y_1 (dry basis) to Y_2 is

$$\int_0^{\theta} dt = \int_{Y_1}^{Y_2} dy/m$$

$$\theta = \int_{Y_1}^{Y_2} dy/m$$

The total mass of water evaporated is $(Y_2 - Y_1)$. Therefore the average drying rate is:

$$\bar{m} = (Y_2 - Y_1) \; / \int_{Y_1}^{Y_2} dy/m$$

Since the experimental results were obtained in terms of the moisture content x, expressed on wet-basis, then,

$$y = x/(1-x)$$

$$dy = 1/(1-x)^2 \, dx$$

$$\bar{m} = [X_2/(1-X_2) - X_1/(1-X_1)] \; / \int_{X_1}^{X_2} dx/m(1-x)^2$$

where X_1 and X_2 are the initial and final moisture contents on wet-basis.

The above integral was evaluated numerically, taking the values of m from the smoothed drying curves.

Drying Tests Without Oil

The drying rates of hog fuel, expressed as kg/h of water, as a function of the moisture content are shown in Figure 9 for heating steam pressure of 200 kPa and in Figure 10 for steam pressure of 250 kPa. One general observation from these results is that the drying rate continuously decreased with the decreasing moisture content, i.e., no constant-rate period was observed. It should be pointed out that the charge moisture content was always higher than 60% and that the results shown here are the ones taken only after the system had reached its operating temperature. Another observation is that the degree of fill affected the drying rates only very marginally; this is somewhat surprising since an increased charge would provide a larger overall particle surface area and, hence, would be expected to result in higher

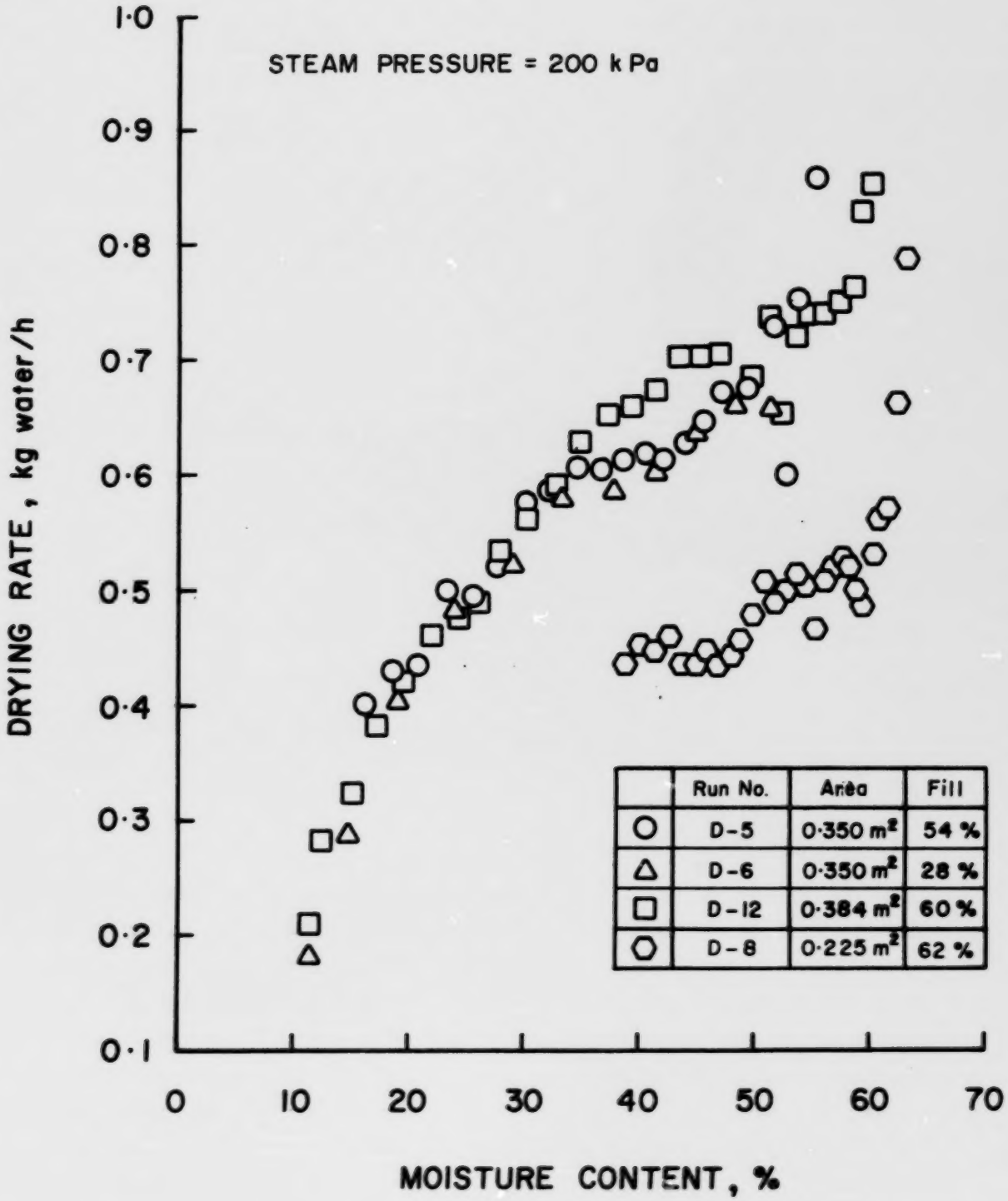


Figure 9. Hog-fuel drying rates as a function of moisture content, at heating steam pressure of 200 kPa.

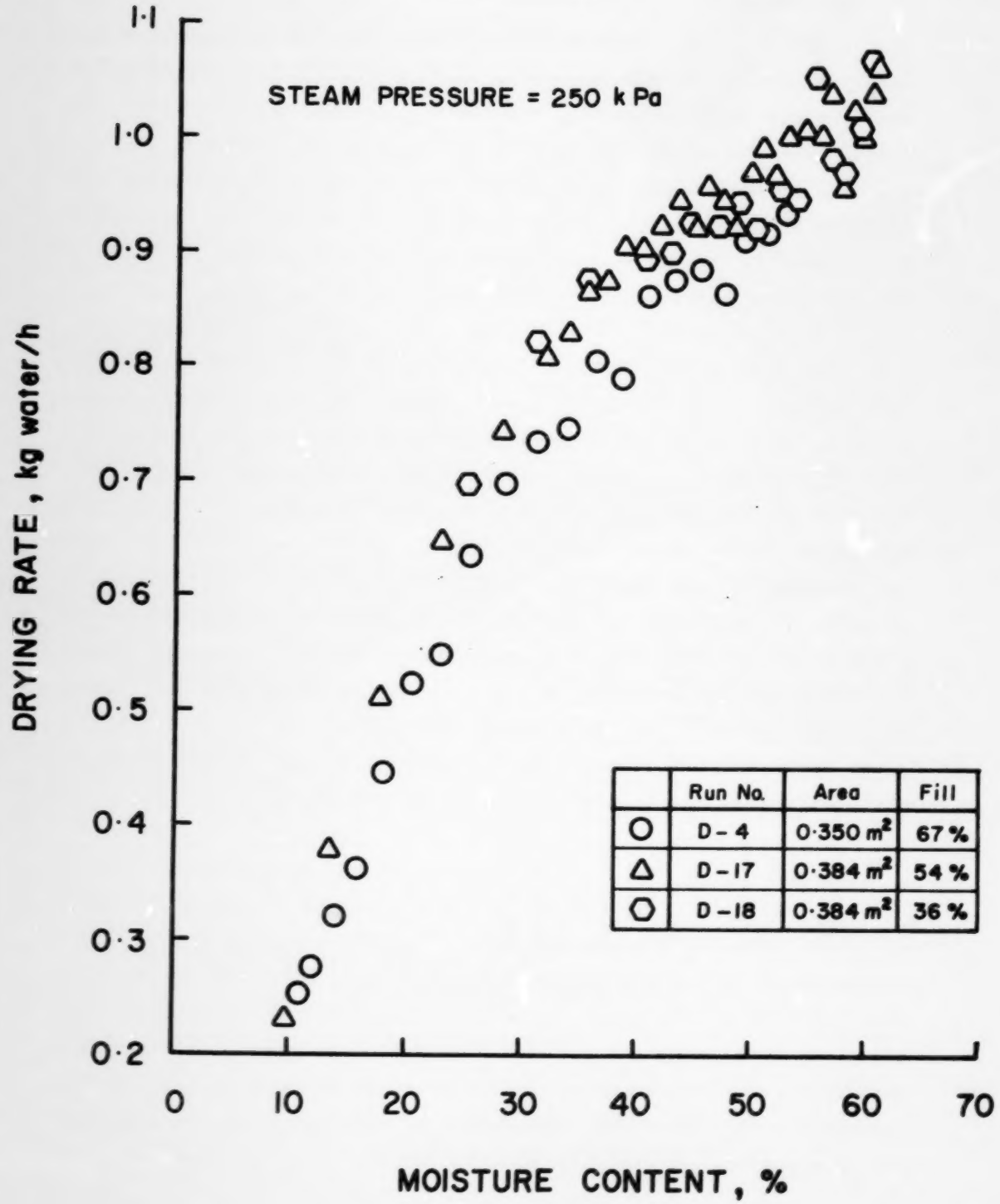


Figure 10. Hog-fuel drying rate as a function of moisture content, at heating steam pressure of 250 kPa.

heat transfer rates. Therefore, it appears that the drying rates were mainly dependent on the condensing surface area, as was confirmed by the fact that when the surface area was reduced by about one-third, the drying rates decreased by proportionate amounts. As expected, increasing the steam pressure from 200 to 250 kPa, with the corresponding increase of the condensing temperature from 120 to 127.5°C, resulted in an increase in the drying rates. The pressure on the hog-fuel side of the vessel was atmospheric in all of the tests.

Using the latent heat of evaporation of water at 100°C, namely, 2260 kJ/kg, the drying rates were converted to heat transfer rates; dividing these values by the temperature differences and the surface area of the tubes gave the overall heat transfer coefficients as a function of the hog-fuel moisture content. The hog-fuel temperature was assumed to be equal to 100°C. Figure 11 shows the results of these calculations for the tests discussed above. It is obvious from Figure 11 that the heat transfer coefficient is primarily a function of the moisture content and that it is essentially insensitive to the operating parameters of the dryer, at least in the range of conditions tested. This suggests that significant improvements in the heat transfer rates would only be possible by fundamental changes in the dryer design, and not by the varying the operating parameters.

The results of the above tests gave overall heat transfer coefficient values ranging between 20 and 70 W/m^2 K with an average value of 55 W/m^2K. This is in agreement with the published data [6] for indirectly heated rotary dryers, which range between 25 and 85 W/m^2K. In a novel pulp dryer [4], where the pulp is conveyed by high-velocity steam along the dryer tubes and the heat is supplied indirectly by high-pressure steam, the heat transfer coefficients are somewhat higher, ranging between 120 and 165 W/m^2K. This would be expected since the particles are much smaller and the pulp/steam velocities are about 30m/s.

For an 800 tpd kraft mill generating about 280 tpd (dry basis) of hog fuel, the heat transfer surface area for a dryer, to reduce the moisture content from 60 to 20%, would be about 10,000 m^2, based on the

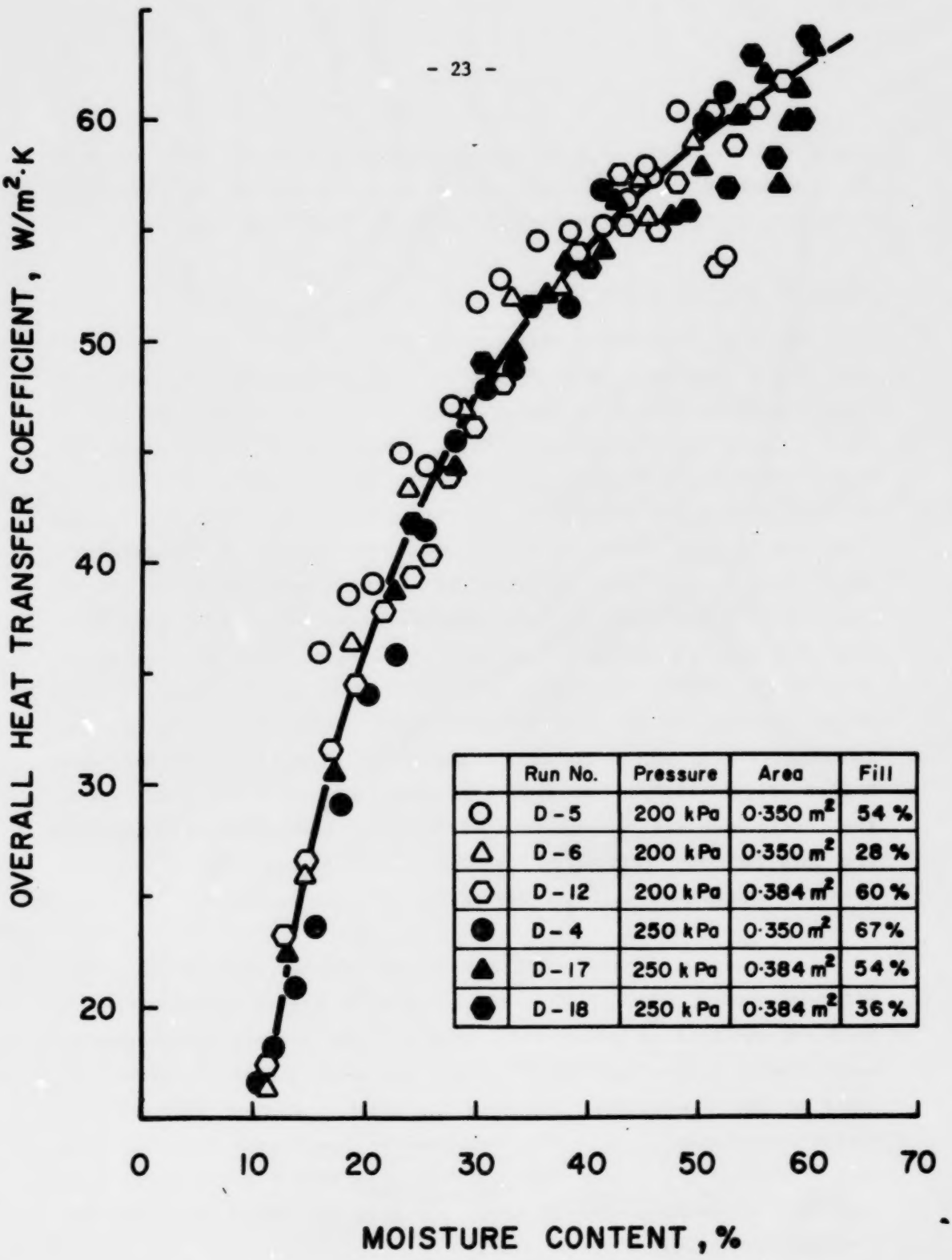

OVERALL HEAT TRANSFER COEFFICIENT, W/m².K
60
50
40
30
20
0
10
20
30
40
50
60
70
MOISTURE CONTENT, %
Run No. | Pressure | Area | Fill
D-5 | 200 kPa | 0·350 m² | 54 %
D-6 | 200 kPa | 0·350 m² | 28 %
D-12 | 200 kPa | 0·384 m² | 60 %
D-4 | 250 kPa | 0·350 m² | 67 %
D-17 | 250 kPa | 0·384 m² | 54 %
D-18 | 250 kPa | 0·384 m² | 36 %

Figure 11. Overall heat transfer coefficient as a function of hog-fuel moisture content, under different dryer operating conditions.

above results. This is a very large surface area, which might prove to be uneconomical. Consequently, for a vapour recompression dryer to be economically feasible, the heat transfer rates must be increased.

Stirred-Oil Drying Tests

Hog fuel from three different sources was used in these tests: (i) from a pulp mill, Mill A, (ii) from a sawmill, Mill B, and (iii) from the PPRIC Chip Debarker. The results of the tests, expressed as heat transfer coefficients (W/m^2K) as a function of the moisture content, are shown in Figures 12 to 14. One general observation that can be made from these results is that the overall heat transfer coefficients were very sensitive to the moisture content of the material, where they dropped from 150 W/m^2K at moisture contents above 60%, to less than 50 W/m^2K at moisture contents below 30%. Figure 12 shows that reducing the amount of solids in the batch, i.e. decreasing the consistency of the oil-solids mixture from 12 to 3%, resulted in a marked increase in the heat transfer rates. This might have been due to better mixing at the low consistency test. Tests with hog fuel from Mill B (Figure 13) did not show a similar increase in the heat flux as the consistency was reduced from 19 to 10%. Tests with material from the Chip Debarker (Figure 14) showed a similar decrease of the heat transfer coefficient with the decreasing moisture content.

A comparison of the results from all of the above tests in the moisture content range of 20 to 60% is shown in Figure 15 (some results were extrapolated to cover this range). The average heat transfer coefficients within the 20-60% range are also listed. Stirred-oil tests at low consistency (3%) gave the highest average heat transfer coefficients, namely 94 W/m^2K, followed by tests with hog fuel from Mill B (87 $W/m^2/K$). Test with hog fuel from Mill A and from the Chip Debarker, at consistencies of about 10% gave the lowest heat transfer coefficients. i.e., about 55 W/m^2K.

The effect of intermittent sampling on the heat transfer rates obtained from the tests can be approximately estimated as follows:

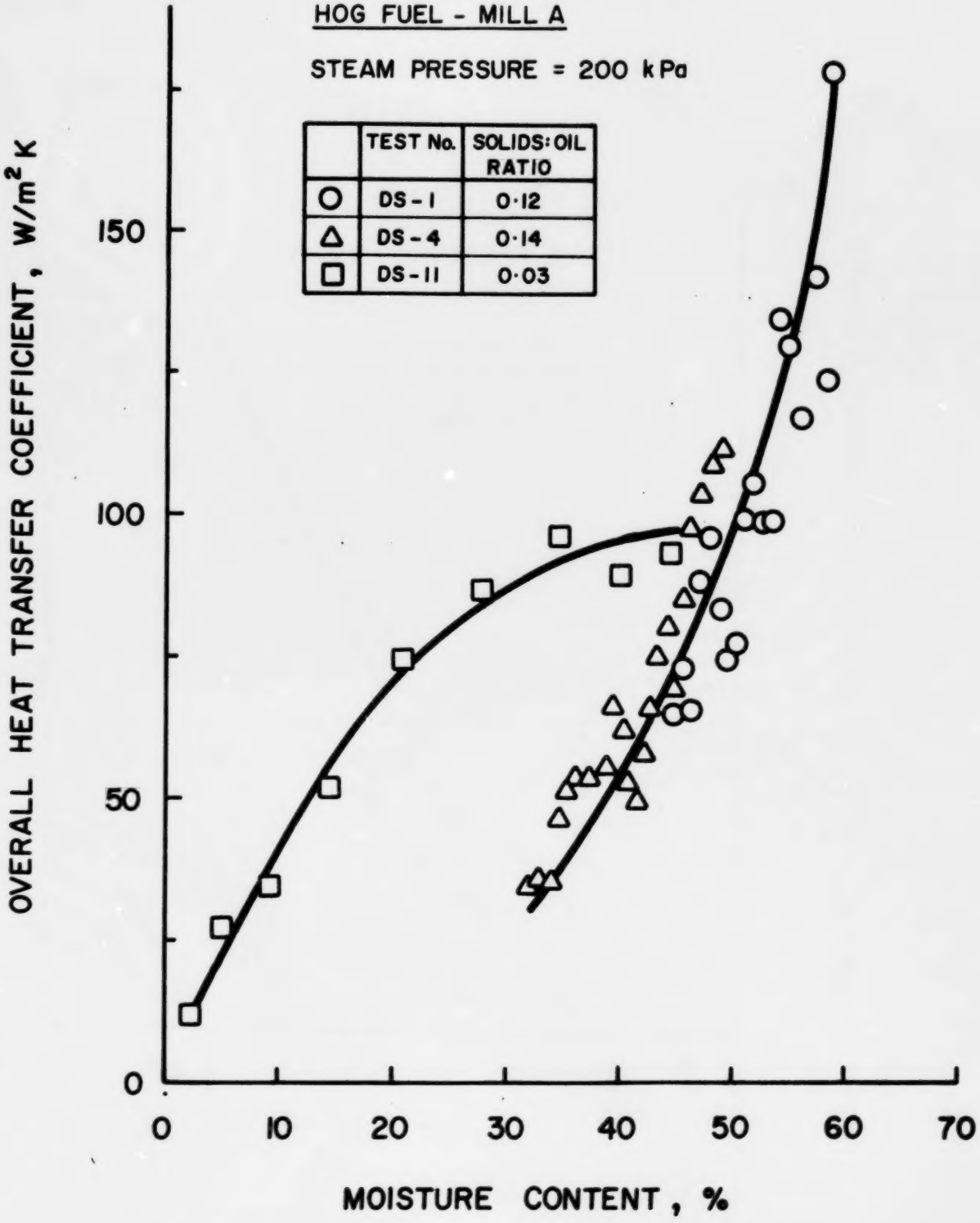


Figure 12. Results of stirred-oil tests on hog fuel from Mill A, under different operating conditions.

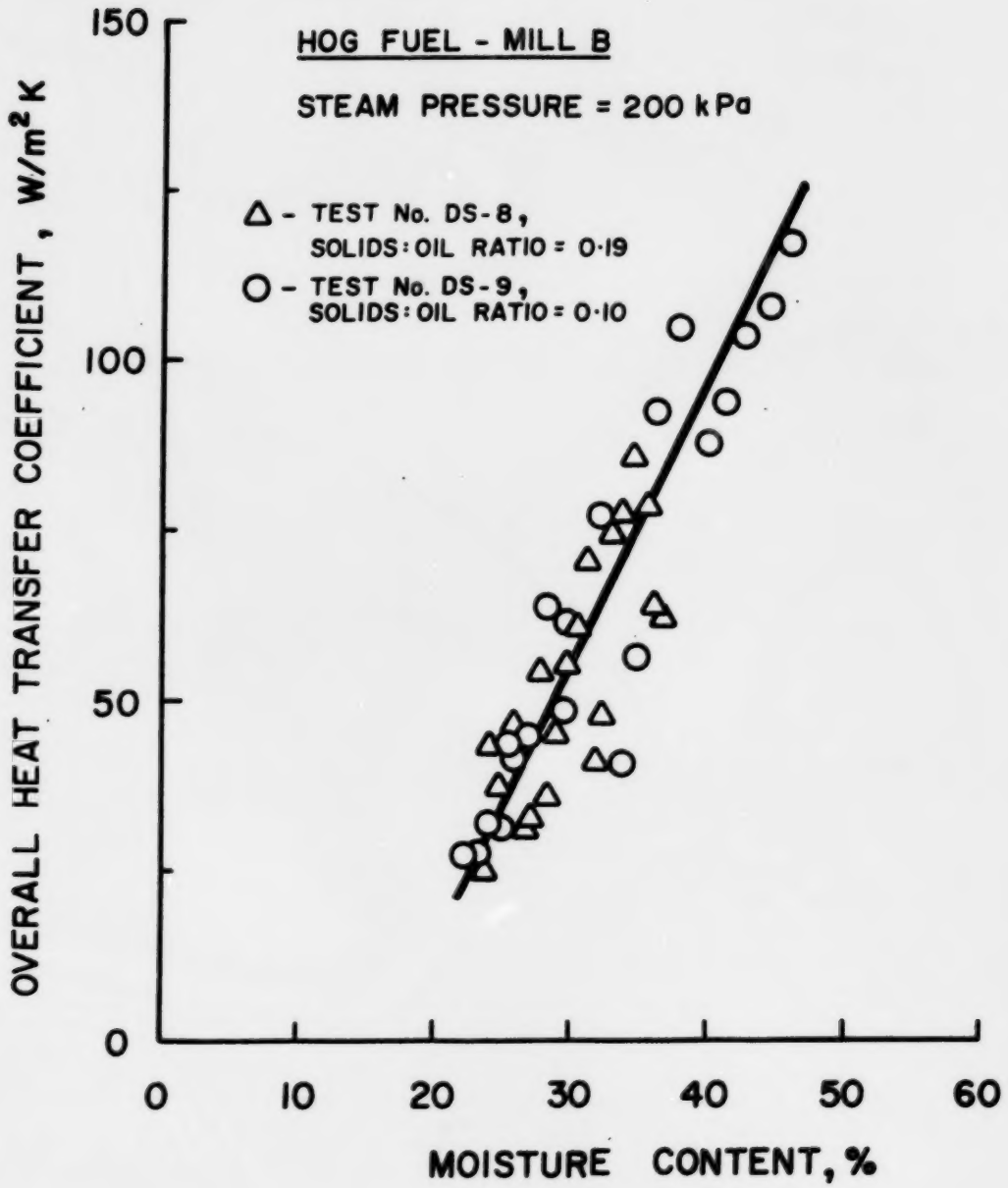


Figure 13. Results of stirred-oil tests on hog-fuel from Mill B, under different operating conditions.

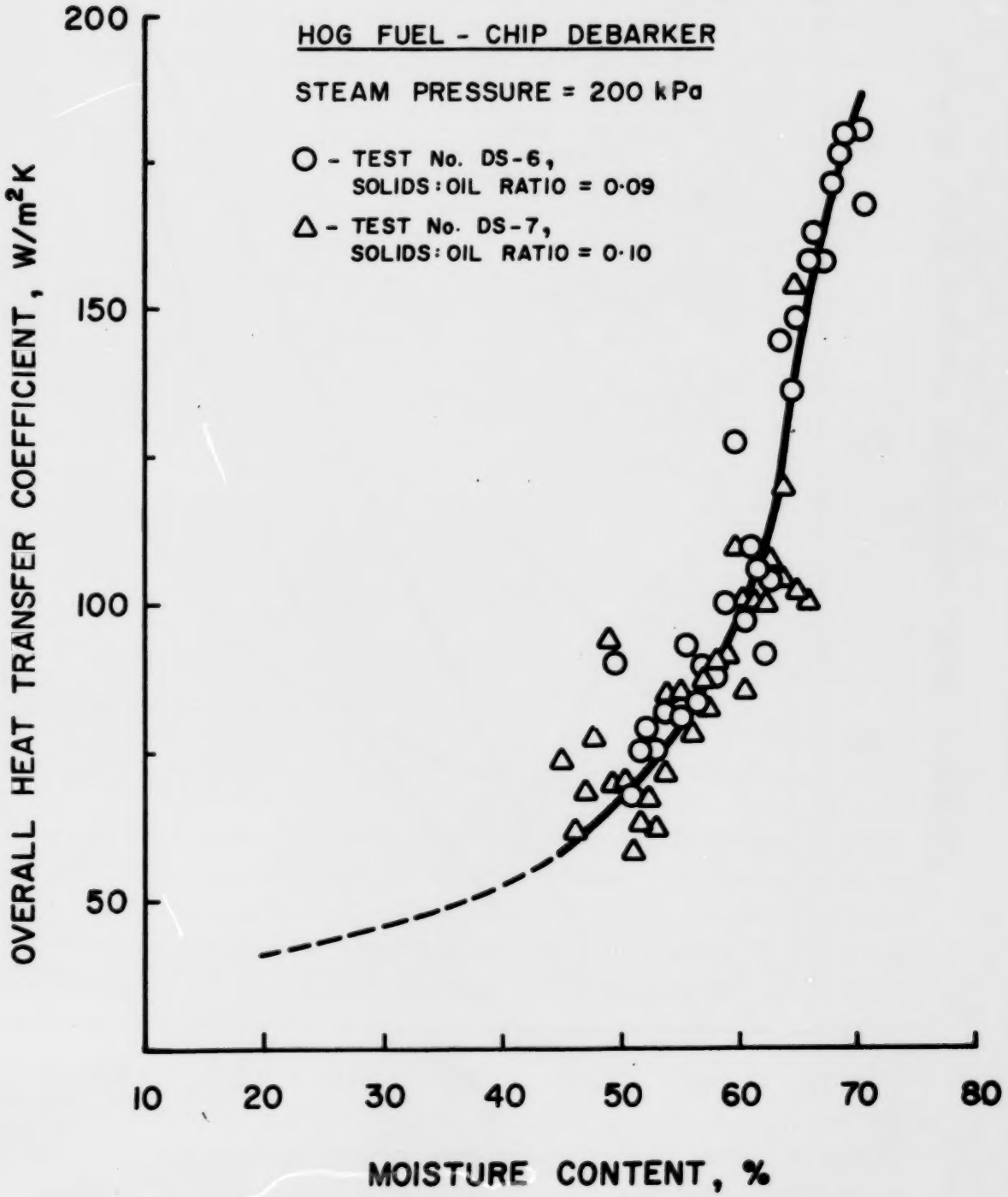


Figure 14. Results of stirred-oil tests on hog fuel from PPRIC Chip Debarker.

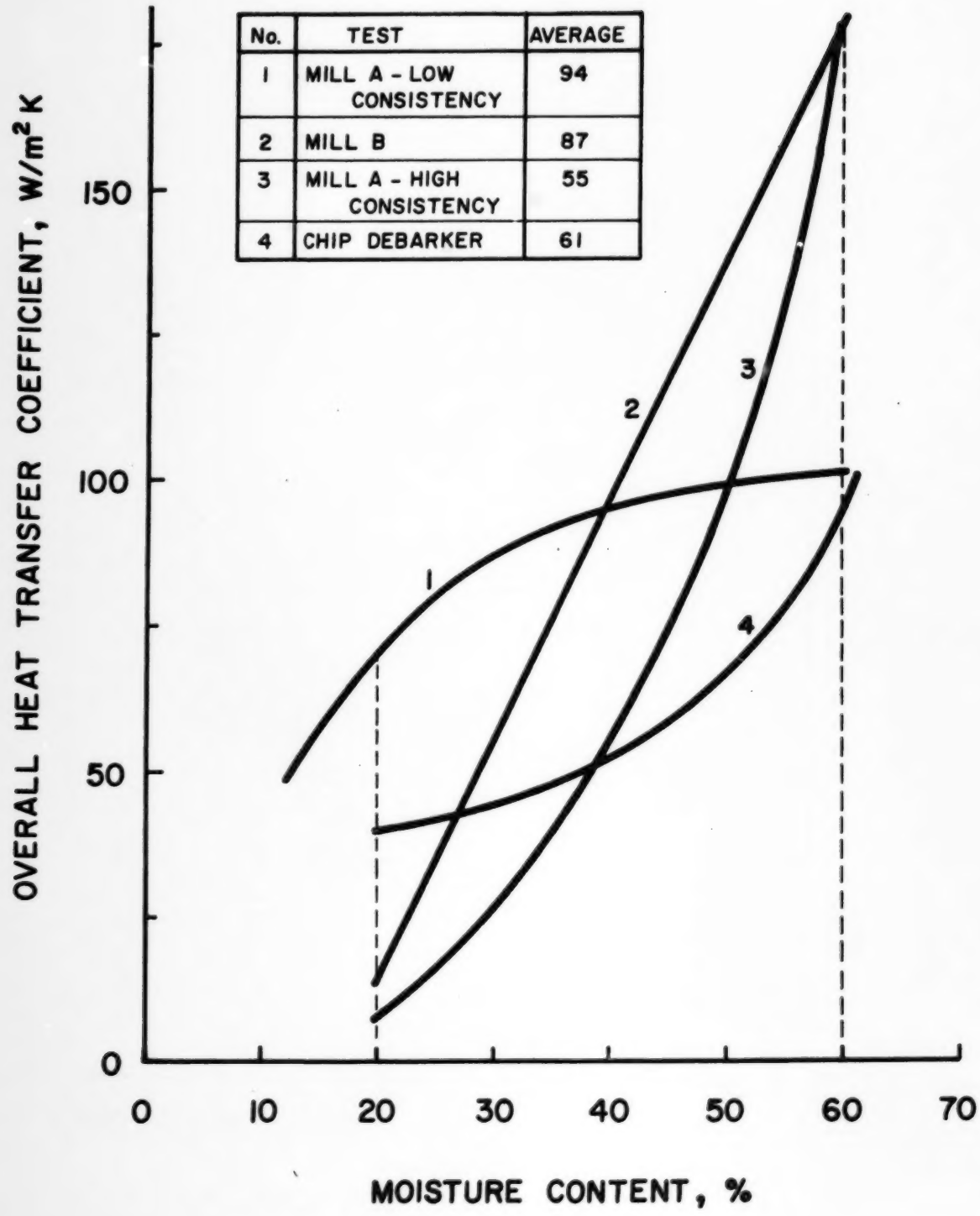


Figure 15. Comparison of results from tests with the rotary batch dryer, using tall oil as an intermediate heat transfer medium.

Assume the mass of condensate collected in the time period Δt is ΔW. Also assume the drying rate during rotation of the vessel is m_1 and during sampling is m_2. Rotation time is Δt_1 and sampling time, Δt_2. Therefore,

$$\Delta W = m_1 \Delta t_1 + m_2 \Delta t_2$$

and

$$\Delta t = \Delta t_1 + \Delta t_2$$

Dividing the first equation by Δt gives,

$$m = \Delta W/\Delta t = m_1 \Delta t_1/\Delta t + m_2 \Delta t_2/\Delta t$$

Usually $\Delta t_1/\Delta t$ ranged between 0.5 and 0.7. Hence

$$m \simeq (0.5 \rightarrow 0.7)\, m_1 + (0.5 \rightarrow 0.3)\, m_2$$

Assuming $m_2 = 0.1\, m_1$, then

$$m \simeq (0.55 \rightarrow 0.73)\, m_1$$

and the drying rate during rotation, m_1, expressed in terms of the average drying rate, m, would be,

$$m_1 \simeq (1.4 \rightarrow 1.8)\, m$$

In other words, on the average, the actual heat transfer rates for a stirred-oil system should be about 60% higher than those listed in Figure 15. This means that the average heat transfer coefficients with stirred-oil should range between 90 and 150 W/m^2K, which are about 1.6 to 2.7 times the drying rates obtained in the tests without the oil.

Circulating-Oil Drying Tests

As mentioned earlier, because of plugging problems with the hog fuel, the circulating oil tests were limited to wood chips. Figure 16 shows the results from a circulating-oil test and from a stirred-oil test (as described in the previous section) using wood chips. In the circulating-oil test the superficial velocity of the oil in the dryer

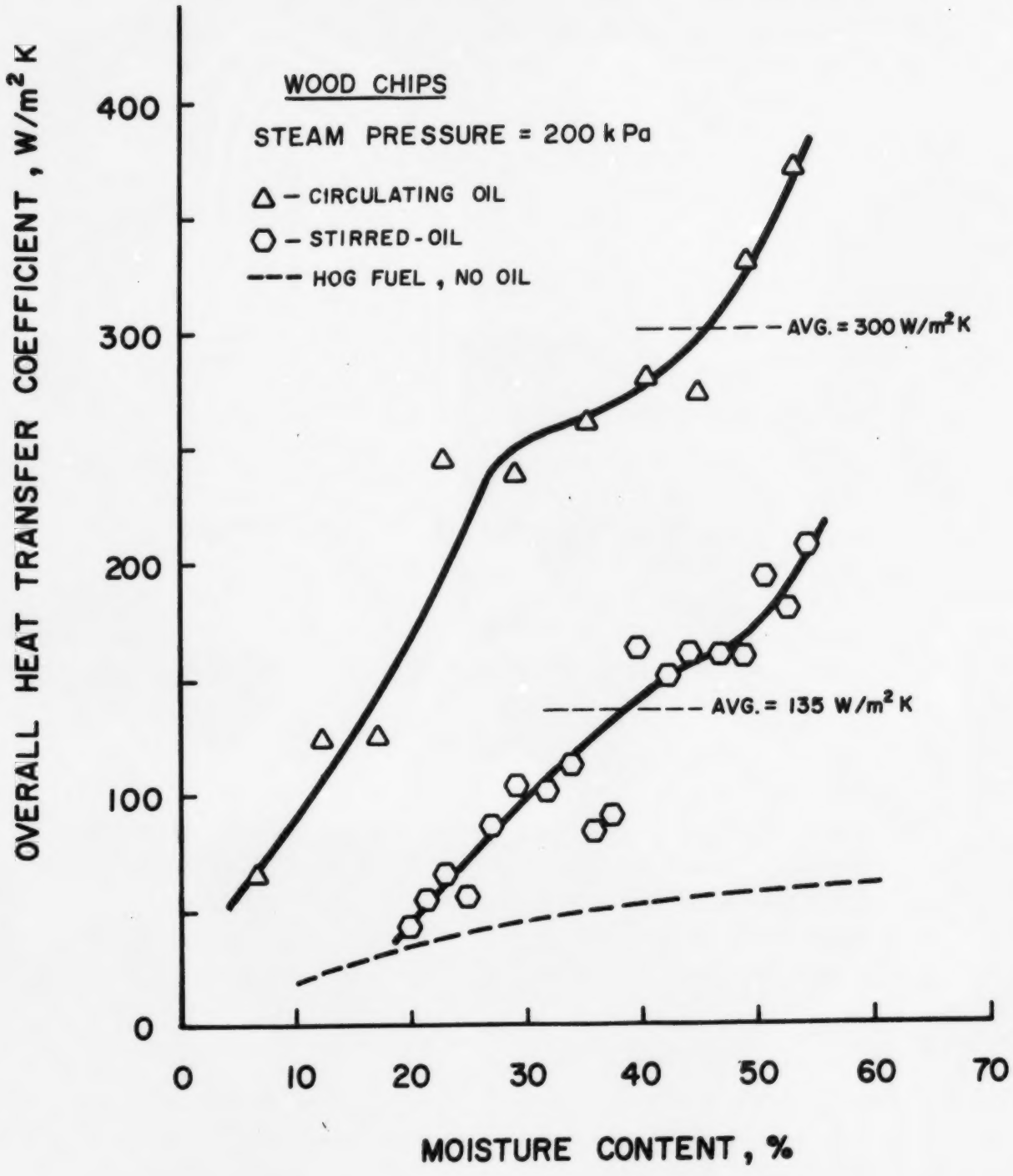


Figure 16. Result of drying wood chips under different operating conditions.

was about 8 cm/s. The average heat transfer coefficients between 60 and 20% moisture was 300 W/m^2K. In the stirred-oil test the average heat transfer coefficient was 135 W/m^2K without correction for intermittent sampling, i.e. about 215 W/m^2K after correction. Hence the circulation of the oil gave drying rates that were 40% higher than those obtained with stirring only. This increase is believed to be due to the higher relative velocities of the oil past both the heat transfer surface and the wood chip particles.

It is worth noting that the heat transfer rates obtained in the stirred-oil dryer with wood chips were 1.3 to 2.5 times the rates obtained with hog fuel, under similar operating conditions. This might have been caused by the difference in the physical nature of the two materials. Wood chips are usually more homogeneous, more rigid and absorb less oil than hog fuel. Consequently, the wood chip particles tend to act as discrete particles in the oil dryer, allowing the oil to move past them at a certain relative velocity. On the other hand, hog-fuel particles tend to absorb significant amounts of oil and form a slurry-like mixture. Therefore, during stirring, very little relative motion may exist between the oil and the hog-fuel particles, and the mixture moves as one, very viscous fluid. Since heat transfer rates are directly proportional to relative motion and inversly proportional to fluid viscosity, heat transfer rates to wood would be higher than those to hog fuel.

CONCLUSIONS

1. Very low moisture contents (< 8%) could be achieved when drying hog fuel indirectly with steam under conditions simulating vapour recompression drying .

2. The rate of drying of hog fuel, without the use of a heat transfer medium, was controlled by the rate of heat transfer from the heat exchanger surface to the water vapour surrounding the hog fuel particles.

3. The overall heat transfer coefficients for drying hog fuel, in absence of a heat transfer medium ranged between 35 and 70 W/m^2K in the moisture content range of 20 to 60% with an average value of 55 W/m^2K. Such drying rates may be too low for an economical process.

4. The use of tall oil as an intermediate heat transfer medium, with the mixing provided by rotating the vessel, gave average overall heat transfer coefficients between 90 and 150 W/m^2K, depending on the source of the hog fuel. As in tests without oil, the heat transfer coefficients decreased significantly with the decreasing moisture content.

5. When oil was circulated past the wet solids, the average overall heat transfer coefficients were even higher, i.e., 300 W/m^2K. However because of plugging problems these tests were limited to wood chips. For comparison, the overall heat transfer coefficient for drying wood chips in the rotating oil system was about 215 W/m^2K. In other words, the circulating-oil dryer gave drying rates about 40% higher than the stirred-oil dryer.

FUTURE PLANS

The next phase in the development of the vapour recompression dryer will be the testing of a pilot-plant set-up with continuous feed and product removal. The design of the system would be based on the following conclusions reached during the course of the present experimental study:

1. Drying rates without the use of a heat transfer medium are too low for an economic process.

2. While increases in the drying rates were achieved by the use of an intermediate heat transfer medium accompanied with stirring, these increases may not be sufficient to improve the economics of the process significantly.

3. Larger increases in the heat transfer rates could be achieved by moving the oil in the heat exchanger at an appreciable velocity (≈1 m/s).

4. Because of the tendency of the hog fuel to compact at any attempt to separate it from the oil, the oil and the hog fuel will have to flow together in the dryer at the same bulk velocity.

A conceptual diagram of the dryer is shown in Figure 17. The dryer would have a vertical shell-and-tube configuration. The wet hog fuel and oil mixture (1-2% consistency) would be pumped up the heat exchanger tubes, which would be heated by the condensation of the re-compressed vapours in the shell. The dried hog-fuel and oil mixture would be separated from the evaporated moisture either in a separate vessel or in an enlarged area at the exit of the exchanger. Some of the mixture would be recycled back to the heat exchanger after the addition of wet hog fuel; the remainder would be passed through a solid/liquid separator, after wnich the oil would be recycled back to the dryer and dried hog fuel pressed and compacted to give high-quality densified fuel. The vapour leaving the dryer would be mixed with a small amount of high-pressure steam and compressed in a mechanical compressor before returning to the heat exchanger shell.

The expected advantages of the proposed configuration would be: (i) the oil, hog fuel and vapour would move together along heat exchanger tubes and, hence, there would be a lesser chance of plugging, (ii) the motion of the oil/solids mixture at a moderate velocity in the heat exchanger tubes would give higher heat transfer rates than those obtained by stirring alone, and (iii) the generation of vapour in the oil/solids stream in the heat exchanger tubes could result in a highly turbulent, foamy two-phase flow which may in turn further improve heat transfer coefficients. The major disadvantage, from heat transfer point of view, is that little relative velocity would exist between the oil and the hog-fuel particles and, hence, heat transfer from the oil to the solids would not be maximized.

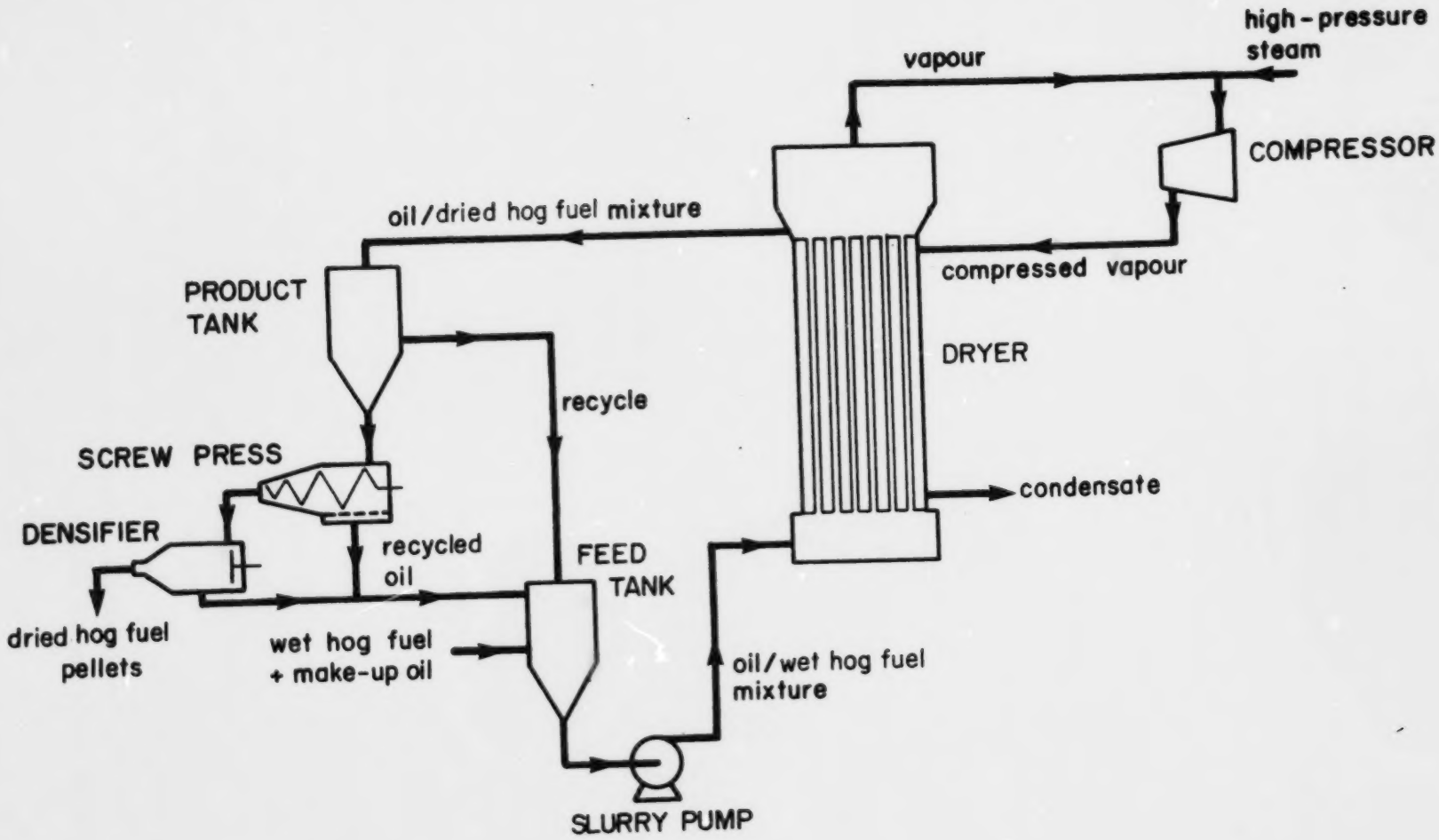


Figure 17. Conceptual diagram of the vapour recompression hog-fuel drying process.

A schematic diagram of pilot-plant set-up to be used in the second phase of the study is shown in Figure 18. The heat exchanger consists of three stainless steel tubes, 50 mm in diameter and 2.5 m long, placed vertically in a steam-heated jacket. The wet hog-fuel/oil mixture would enter at the bottom of the tubes and the vapour/dried hog-fuel/oil stream would leave from the top. The vapour would be separated in a flash tank, containing demisting devices, before condensation and collection. The dried solid/oil mixture would go into a settling tank where part of the solids would be removed via a screw press. The remaining oil and dried solids would be recycled to the feed tank where they would be mixed with fresh wet hog fuel and make-up oil. The mixture from this tank would be conveyed to the heat exchanger with a slurry pump.

The surface area of the exchanger would be about 1.1 m^2. Assuming heat flux between 6 and 10 kW/m^2, the drying rates would be 9 to 15 kg/hr of water which corresponds to 7 to 10 kg/hr hog fuel (dry basis).

PRELIMINARY ECONOMICS

When hog fuel is dried using a vapour recompression dryer, its useful heating value increases from 12.5 MJ/kg (dry basis) at 60% moisture to 15.4 MJ/kg (dry basis) at 20% moisture (Appendix I). The electricity required for vapour recompression is 0.3 MJ/kg hog fuel. If oil is used to enhance heat transfer some will be entrained in the hog fuel to the extent of 5% of the mass of the fuel (after densification). Including this, the heating value of the dried fuel increases to 17.2 MJ/kg (dry basis).

Example: A 50-MW hog-fuel boiler produces only 30 MW of steam because of the high moisture content of the fuel. Firing dried fuel would allow the boiler to achieve its rated capacity. Therefore, some oil would be saved in other boilers in the mill.

Steam requirements = 50 MW

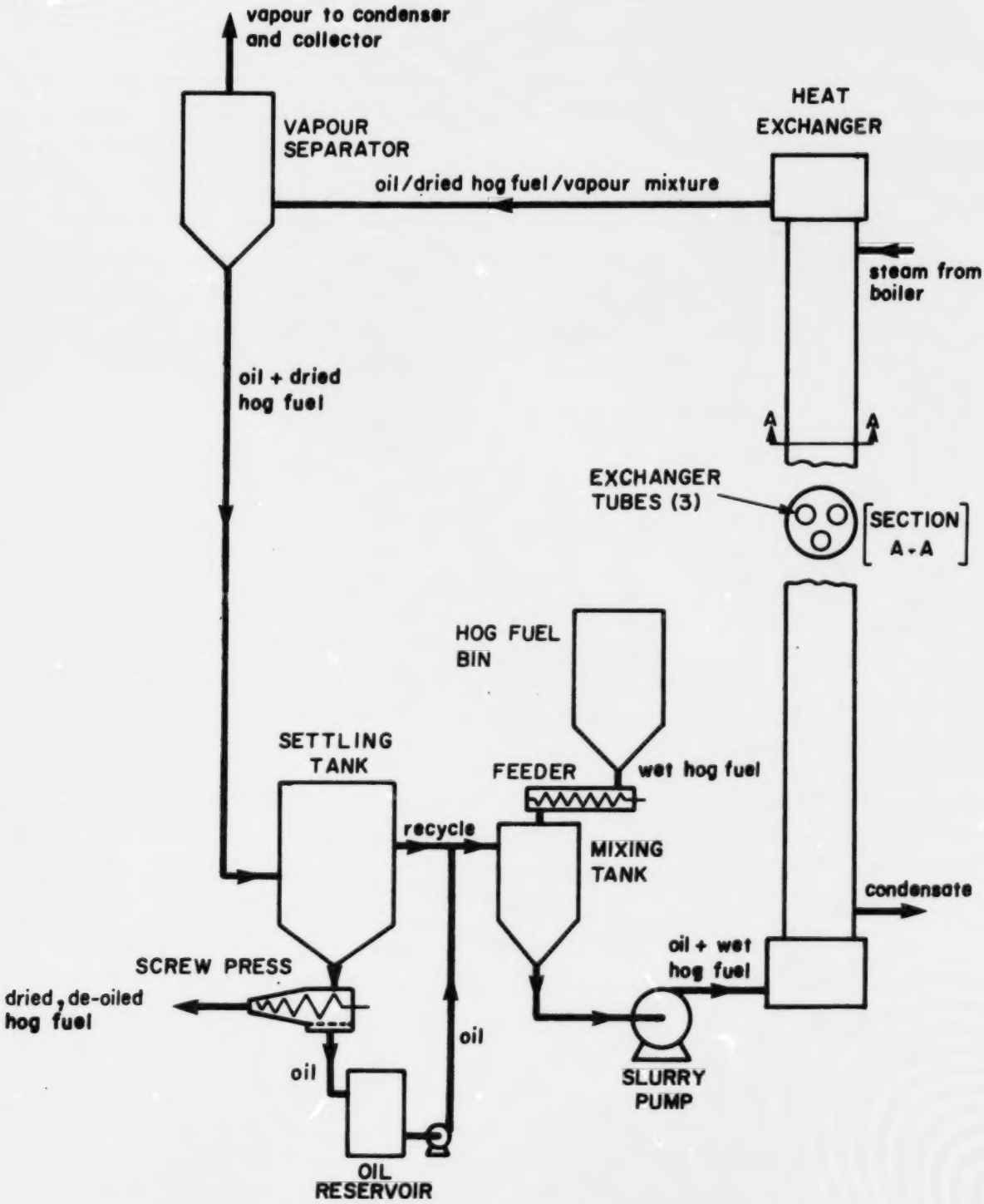


Figure 18. Schematic diagram of the proposed pilot scale experimental set-up.

When firing wet hog fuel : 205 tpd hog fuel produces 30 MW
plus oil : 300 bbl/d oil produces 20 MW

When firing dried fuel : 250 tpd hog fuel produces 50 MW
(oil entrained in hog fuel = 80 bbl/d)

Increase in hog-fuel consumption = 250-205 = 45 tpd
Decrease in oil consumption = 300- 80 = 220 bbl/d

Hog-fuel dryer capacity required = 250 tpd (dry basis)
Dryer heat load = 3.4 MJ/kg ≡ 9.8 MW (see Appendix II)

Using the heat transfer coefficients obtained with the circulating oil system (300 W/m^2k) and an overall temperature difference of 20°C:

Surface area required = 9800 kW/(0.3 x 20) = 1640 m^2

(The heat transfer coefficients are expected to be higher for the proposed configuration which is very similar to vertical tube evaporators, e.g., in the evaporation of kraft black liquor overall heat transfer coefficients range between 1.0 and 1.7 kW/m^2K).

Compressor load = 0.3 MJ/kg ≡ 870 kW

Capital Costs

It is difficult to estimate the cost of each item of the dryer because of the many uncertainties regarding the actual design of the system. However the heat exchanger/compressor part is very similar to standard vapour recompression black liquor evaporators. Hough [7] estimated the direct cost (materials & labour) of an evaporator system with a capacity of 136,000 kg/hr water evaporated to be U.S. $1.8 million (1977). The elecrical consumption of the system was 3900 kW. Based on an estimated heat flux of 16 kW/m^2, the heat transfer surface area would be about 5600 m^2. Both the heat transfer surface and the compressor are about four times the size of those required for the hog-fuel dryer. Assuming,

$$\text{Cost} = \text{Constant} \times (\text{size})^{0.7}$$

then the direct cost of the heat exchanger/compressor part of the hog-fuel dryer would be,

$$\text{direct costs} = \$1.8 \times 10^6 \times (1/4)^{0.7}$$

$$\text{direct costs} = \$700,000$$

However, these figures are given in 1977 U.S. dollars. Inflation between 1977 and 1982 is estimated at 60%. The Canadian dollar is equivalent to $0.8 U.S. Therefore,

$$\text{direct costs} = \$1.6 \times 700,000/0.8 = \text{Can. } \$1.4 \text{ million}$$

Indirect costs and contingency are typically 60% of direct costs, hence,

$$\text{Installed costs} = 1.4 \times 1.6 \times 10^6 = \$2.24 \text{ million (1982)}$$

Other major items in the dryer are (i) feeding system, (ii) tanks and settlers and (iii) presses and compactors. The purchased costs of these items are estimated to be the following:

Hog-fuel feeding system	= $ 100,000
Tanks and settlers	= $ 100,000
Presses and pelletizer	= $ 500,000
Total purchased costs	$ 700,000

Total direct costs (labour & material) are typically 2.5 times the purchased costs, and indirect costs 0.6 times direct costs. Hence,

$$\text{Installed costs} = 2.5 \times 1.6 \times \$700,000 = \$2.8 \text{ million.}$$

Therefore, the total installed costs of the dryers would be,

$$\text{Total capital investment} = \$2.8 + 2.24 \text{ million}$$

$$\text{T.C.I.} = \$5.04 \text{ million}$$

Overall Economic Assessment

Oil savings at $26/bbl	= $2,002,000	(1)
Hog fuel at $20/t (dry basis)	= $ 315,000	(2)
Savings in fuel costs [(1)-(2)]	= $1,687,000	(3)
Labour, 1 man-yr at $17/h	= $ 143,000	(4)
Electricty, 1000 kW at 2c/kWh	= $ 168,000	(5)
Maintenance, 5% T.C.I.	= $ 250,000	(6)
Operating costs [(4)+(5)+(6)]	= $ 561,000	(7)
Gross revenue [(3) - (7)]	= $1,126,000	(8)
Capital charges, 15% investment	= $ 756,000	(9)
Taxable income [(8)-(9)]	= $ 370,000	(10)
Taxes, 50% income	= $ 185,000	(11)
Cash flow [(8)-(11)]	= $ 941,000	(12)

Engineering return on investment (R.O.I.) = Cash Flow/T.C.I.

R.O.I. = $941,000/$5,040,000 = 18.7%

Assuming plant life of 15 years, the discounted cash flow rate of return is,

D.C.F. ROR = 16.5% after tax

These returns would be even higher as the price of fossil fuel increases.

ACKNOWLEDGEMENTS

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P. Biocca

D. Callender

T. Marra

REFERENCES

1. Johnson, R.C., 'Some Aspects of Wood Waste Preparation for Use as a Fuel', TAPPI 58 (7), 102 (1975).

2. McDermott, R.J., Morris, K.W. and Sanderson, J.G. 'A Better Boiler for Wet Hog Fuel', paper presented at the 66th Annual Metting, Technical Section, CPPA, Montreal, Jan. 29 to Feb. 1, 1980.

3. Svensson, C., 'Back Pressure Drying, A New System for Hogged Fuels', Svensk Papperstidning 82 (10), 281 (1979).

4. Svensson, C., 'Novel Pulp Dryer Promises Energy Savings', Pulp Pap. Int'l., 93 (June, 1980).

5. MacCallum, C., Blackwell, B.R. and Torsein, L., 'Cost Benefit Analysis of Systems Using Flue Gas or Steam for Drying Wood Waste Feedstocks', ENFOR Report, Project C-96, May 31, 1981.

6. Perry, R.H. and Chilton, C.H. (eds.), Chemical Engineers' Handbook, 5th ed., McGraw-Hill, 1973, p. 20-41.

7. Hough, G.W., 'Pre-evaporation of Kraft Black Liquor by Vapour Recompression Evaporation', TAPPI 61 (7) (1978).

APPENDIX I

VAPOUR RECOMPRESSION DRYER MASS AND ENERGY BALANCES

The mass and energy balances are carried out for the process shown in Figure 1, i.e., when no oil is used. It is also assumed that no flue gas is used. The use of flue gas would, of course, increase the energy savings of the process. However, there is a certain cost associated with installing a flue gas recovery system which is very mill-specific and, hence, difficult to evaluate at this point.

ASSUMPTIONS

Initial wood moisture content	=	60%
Final wood moisture content	=	20%
Wood inlet temperature	=	25°C
High pressure steam : pressure	=	1.15 MPa
temperature	=	186°C
enthalpy	=	2783 kJ/kg (0°C reference temperature)
Compressor : inlet pressure	=	100 kPa
outlet pressure	=	200 kPa
adiabatic efficiency	=	70%
motor efficiency	=	90%
Heat capacity : Water	=	4.2 kJ/kg K
Wood	=	1.4 kJ/kg K
Steam	=	2.0 kJ/kg K

CALCULATIONS

Basis	= 1 kg dry wood	
Water in with wood	= 1 x 0.6/(1-0.6)	= 1.5 kg
Water out with wood	= 1 x 0.2/(1-0.2)	= 0.25 kg
Water evaporated	= 1.5 - 0.25	= 1.25 kg

Compressor

To prevent steam condensation in the compressor, high-pressure steam is injected to superheat the vapour to 105°C.

Steam enthalpy (saturated at 100 kPa) = 2676 kJ/kg

Steam enthalpy (superheated to 105°C, at 100 kPa) = 2690 kJ/kg

Heat required for superheat	= 1.25 x (2690-2676)	= 17.5 kJ
High pressure steam required	= 17.5 / (2783-2687)	= 0.18 kg
Total steam flow to compressor	= 1.25 + 0.18	= 1.43 kg

Assuming isentropic compressiom, the enthalpy of the steam leaving the compressor, H_{ad}, is:

$$H_{ad} = 2820 \text{ kJ/kg}$$

or

$$\Delta H_{ad} = 2820 - 2687 = 133 \text{ kJ/kg}$$

$$\Delta H_{actual} = \Delta H_{ad}/\text{adiabatic efficiency} = 133/0.7 = 190 \text{ kJ/kg}$$

Theoretical compressor work = 1.43 x 190 = 272 kJ

Electrical consumption = theoretical work/motor efficiency
= 272/0.9 = 302 kJ

Enthalpy of steam leaving compressor = 2690 + 190 = 2880 kJ/kg

Dryer

Shell-side

$$H_{in} = [H_{wood} + H_{water}]\ 25°C$$

$$H_{in} = (1 \times 1.4 + 1.5 \times 4.2) \times (25 - 0) = 193 \text{ kJ}$$

$$H_{out} = [H_{wood} + H_{moisture} + H_{vapour}]\ 100°C$$

$$H_{out} = (1 \times 1.4 + 0.25 \times 4.2) \times (100-0) + 1.25 \times 2676 = 3590 \text{ kJ}$$

$$\Delta H_{shell-side} = 3590-193 = 3397 \text{ kJ}$$

Condenser-side

$$H_{condensate} = 4.2 \times (120-0) = 504 \text{ kJ/kg at } 120°C \text{ (200 kPa)}$$

$$\Delta H_{steam} = 1.43 \times (2880-504) = 3398 \text{ kJ}$$

which is equal to the shell-side requirements.

High pressure steam required = 0.18 kg
Heat contained in h-p steam = 2783 - 4.2 x (25-0) = 2678 kJ/kg
Total heat in steam = 2678 x 0.18 = 482 kg

In other words, 302 kJ of electricity to operate the compressor plus 482 kJ of steam from the boiler would be required.

APPENDIX II

THERMODYNAMICS OF WOOD COMBUSTION

Wood Composition

C = 50%
H = 6%
O = 43%
Ash = 1%

Basis of Calculation = 1 kg dry wood

Wood Combustion

$$C + O_2 \rightarrow CO_2$$

Moles of C = 0.5/12 = 0.0417 kmol

$$H_2 + \tfrac{1}{2} O_2 \rightarrow H_2O$$

Moles of H_2 = 0.06/2 = 0.03 kmol

Oxygen required for combustion = 0.0417 + 0.03/2 = 0.0567 kmol

Oxygen in with wood = 0.43/32 = 0.0134 kmol

Oxygen to be supplied by combustion air = 0.0567-0.0134 = 0.0433 kmol

Air required = 0.0433 x 32 + 0.0433 (0.79/0.21) x 28 = 5.937 kg

Assume 30% excess air is required

Total air in = 1.3 x 5.937 = 7.72 kg

Combustion Gas Composition

CO_2 = 0.0417
H_2O = 0.03 + moisture in wood
O_2 = 0.013
N_2 = 0.2114

Heat Effects

(The reference temperature in these calculations is 25°C)

$$\text{Wood} + O_2 \rightarrow CO_2 + H_2O_{(l)}$$

$\Delta H_{combustion}$ = 19 MJ/kg dry wood at 25°C

$\Delta H_{vaporiz}$ = 2.441 MJ/kg water

Water produced by combustion = 0.03 kmol = 0.54 kg/kg wood

$\Delta H_{vaporiz}$ = 0.54 x 2.441 = 1.32 MJ/kg wood

Net $\Delta H_{combustion}$ = 19 - 1.32 = 17.68 MJ/kg dry wood

Boiler Heat Balance

Assume hog fuel enters boiler at m moisture content and that gases leave at 150°C.

Heat capacity of gases between 25 and 150°C =

CO_2 : 40 kJ/kmol K
H_2O : 34 kJ/kmol K
O_2 : 30 kJ/kmol K
N_2 : 29 kJ/kmol K

Basis: 1 kg dry wood

Moisture in = m/(1-m) (kg)

Combustion gas:

CO_2 = 0.0417 kmol
N_2 = 0.02114
O_2 = 0.013
H_2O = 0.03 + m/18(1-m)

Enthalpy of flue gas = $\left(0.0417 \times 40 + 0.2114 \times 29 + 0.013 \times 30 + 34 \times (0.03 + m/18(1-m))\right)$ (150-25)

$$\Delta H_{f.g.} = 1.151 + 0.236\ m/(1-m) \quad \text{(MJ)}$$

Heat required to evaporate moisture = 2.441 m/(1-m) (MJ)

Useful heat in boiler = 17.68 - (1.151 + 0.236m/(1-m)) - 2.441m/(1-m)

$$Q_{boiler} = 16.53 - 2.677\ m/(1-m) \quad \text{(MJ)}$$

Operation Without Dryer

Assume initial moisture content of hog fuel, m = 60%

$$Q_{boiler} = 16.53 - 2.677 \times 0.6\ /(1-0.6)$$

$$Q_{boiler} = 12.5 \text{ MJ/kg hog fuel (dry basis)}$$

Overall efficiency = 12.5/19 = 66%

Operation With FLue-Gas Dryer

The dryer-boiler system can be considered as one unit, except that the flue gas leaves at 100°C instead of 150°C. It is immaterial at what temperature the gases enter the dryer for this affects the kinetics of the drying only.

Enthalpy of flue gas leaving dryer = $\left(0.0417 \times 40 + 0.2114 \times 29 + 0.013 \times 30 + 34 \times (0.03 + m/18(1-m))\right) \times (100-25)$

$$\Delta H_{f.g.} = 0.69 + 0.142\ m/(1-m)$$

$$Q_{boiler} = 17.8 - (0.69 + 0.142\ m/(1-m)) - 2.441\ m/(1-m)$$

$$Q_{boiler} = 16.99 - 2.583\ m/(1-m)$$

At m = 60% Q_{boiler} = 13.1 MJ /kg hog fuel (dry basis)

Overall efficiency = 13.1/19 = 69%

Operation With Vapour Recompression Dryer

Assume hog fuel leaves dryer at 20% moisture. Since the flue gases leave the system at 150°C, then:

$$Q_{boiler} = 16.53 - 2.677\ m/(1-m)$$

For m = 20%, Q_{boiler} = 15.86 MJ

This is not the net heat generated in the boiler since the operation of the vapour recompression dryer requires some high-pressure steam. This was calculated in Appendix I and is equivalent to 0.48 MJ/kg hog fuel.

Therefore:

Net Q_{boiler} = 15.86 - 0.48 = 15.38

Net efficiency = 15.38/19 = 81%

It should be pointed out that 0.3 MJ/kg hog fuel (dry basis) would also be required as electricity to operate the compressor.

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